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HIGH-VELOCITY, HIGH-EFFICIENCY AEROSOL FILTRATION

David Leith, et al

Harvard School of Public Health

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16. ABSTRACT
The report gives results of bench- and pilot-scale studies of the dust collection characteristics of fabric filters. Techniques for measuring dust deposit porosity as a function of cloth characteristics and filtration velocity on a bench-scale filter have been developed and are described. A method for impregnating and slicing the dust deposit for examination under the electron microscope is also described. For the pulsejet pilot-scale filter, flyash penetration decreased as the dust deposit thickened, increased with increasing filtration velocity, and remained relatively constant for particles down to 0.30 micrometers diameter. Three dust emission mechanisms were investigated, using chemically tagged flyash. Penetration by straight-through dust loss falls off rapidly after cleaning, but later increases. Seepage of dust through the fabric was constant throughout the filtration cycle. Dust lost as pinhole plugs increased after cleaning, but later declined; however, the pinholes may open the way for further emission by the straight-through mechanism. Fabric cleaning was a problem in both the pulsejet and shaker cleaned filters during high velocity operation; redesign of commercial equipment is necessary.

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by

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SECTION I

CONCLUSIONS

Flyash penetration through a pulse-jet fabric filter increases with filtration velocity, decreases with dust accumulation after cleaning, but remains relatively constant for all particles down to 0.30 μm diameter. Immediately after cleaning, most dust passes through by the "straight through" mechanism. However, "seepage" and "pinhole plug loss" mechanisms rapidly rise in importance. Pinholes in the dust deposit provide attractive sites for emission by the "straight through" mechanism.

Techniques have been developed to permit examination of dust cake intrastructure and internal geometry. Compression of the cake structure occurs after deposition. Cake porosity is greatly affected by filtration velocity, even when hard, spheroidal particles are used. Loss of deposited dust by pinhole plugs was observed, but could be eliminated by modification of the cake substrate.

Pulse-jet and shaken bag cleaned filters both increased in penetration with increasing filtration velocity and pressure drop. Substantial changes in filter configuration are necessary before the full potential of high velocity filtration can be realized.

SECTION II

RECOMMENDATIONS

As a result of this program, many promising new techniques have been developed for characterizing the operation of fabric filters. Statistical procedures were developed for the estimation of filter efficiency confidence intervals. Confidence intervals are essential for the meaningful interpretation of dust collector efficiency data. A technique for tagging dusts with a variety of tracer substances was developed that made it possible to determine the mechanisms by which dust passed through the filter.

Methods of dust deposit examination were determined, which allow estimation of internal cake geometry, including cake porosity. Needed modifications to commercial fabric filtration equipment have been identified, which will permit modification for high velocity operation.

At present, these new and powerful techniques and this design information have only been applied to a limited number of fabrics and dusts. To make effective use of the development effort already made, a substantially augmented program of fabric filter study should be undertaken.

This further study would use the new techniques described here to study a range of dusts, fabrics, cleaning methods, and operating conditions including conventional and high velocity filtration, which is indicative of industrial practice. In this way, the new techniques can have most effect on improving the performance of fabric filters in the field, and can best stimulate the development of a new generation of high velocity-high efficiency filters.

SECTION III

INTRODUCTION

This is a final report for Research Grant R801399 entitled High Velocity - High Efficiency Aerosol Filtration conducted from April 1973 to April 1975 at the Harvard Air Cleaning Laboratory, a research group within the Department of Environmental Health Sciences. Three distinct but interrelated programs were carried out under this general heading during the grant period, making it desirable to prepare this final report in three sections.

Section IV, "PARTICLE COLLECTION IN A PULSE-JET FABRIC FILTER" is a condensation of a thesis submitted by David Leith as partial fulfillment of the requirements for the granting of the degree of Doctor of Science by Harvard University. A shorter version of this thesis was presented at the Annual Meeting of the Air Pollution Control Association in Boston, June 1975 and has been submitted for publication in the Journal of the Air Pollution Control Association.

In the course of this study, mass penetration and particle size penetration data were collected for pulse jet bag filters,and a filtration model for this type of fabric filter was developed that covered filtration velocity, particle size, penetration by mass and by size, and the cleaning-filtering cycle (i.e., deposit thickness). A technique for tagging dusts with a variety of tracer substances was developed that made it possible to evaluate the fraction of penetrating dust that was associated with straight through passage, seepage through the cake and cloth, and pinhole plug loss. This was accomplished by adding successive layers of differently tagged, aerosolized fly ash to the filter and cake over a complete filtration cycle and analyzing the penetrating dust for the tracer materials added in known sequence. This is a very promising and a powerful technique for examining the structure of filter cakes and their failure modes, and it can be applied profitably to all types of cloth filters and all types of aerosols.

Section V, "HIGH VELOCITY CAKE FILTRATION" was part of a doctoral thesis research project conducted by Stephen Rudnick. Oral presentation and Journal publication have been postponed pending the development of additional experimental data needed to evaluate empirical constants in the design equations.

This work was concerned with the infrastructure of filter cake formation and its internal geometry utilizing Arizona fine test dust as the principal cake material. Techniques and apparatus were developed to produce filter cakes at constant mass air flow rates (i.e., compensating for filter resistance buildup) and with differently sized fractions of the test dust; utilizing air elutriation methods for size stratifying a dynamic aerosol cloud. Cake structure was "frozen" by careful infiltration with a thermosetting plastic monomer followed by polymerization. After solidification, the dust deposit can be treated like a mineral specimen, i.e., it can be sliced, polished, and examined under the scanning electron microscope. These studies showed that there is compression of the cake structure after deposition when the deposit is subjected to high pressure and that cake porosity is greatly affected by filtration velocity. These observations apply to a mineral dust of spheroidal shape and the effects of velocity and pressure may become much more marked for fluffy agglomerated particles such as a freshly formed zinc oxide fume. Pinhole plug losses were easily observed by this operational mode and, under the microscope, were seen to have a diameter of 200 - 300 µm. Pinhole punctures could be eliminated by using two layers of cloth supported by a wire screen and by low filtration velocities. This suggests that pinhole punctures result from movement of the cloth substrate as a result of flexure under pressure changes, rather than from failure in the dust cake structure.

Section VI , "APPLICABILITY OF CONVENTIONAL BAGHOUSES TO HIGH VELOCITY OPERATION describes a number of high velocity filtration studies (to a filter pressure drop of approximately 40 in. w.g.) conducted on pulse-jet and shaking bag filters to examine important filtration performance factors. Penetration showed a strong tendency to increase with increasing filtration velocity and became unacceptable at the highest filtration velocities for the pulse-jet filter. Stretching of the fabric with opening of pores and destruction of the integrity of the filter cake are suspected. It has been concluded on the basis of these observations that future work on high velocity fabric filtration should not be conducted with conventional designs as these are clearly unadaptable to the very different design requirements of high velocity-high pressure drop fabric filtration.

Section IV

PARTICLE COLLECTION BY A PULSE-JET FABRIC FILTER

GENERAL

The industrial gas cleaning community recognizes fabric filters as the equipment best able to achieve high particle collection efficiency over the entire spectrum of particle sizes. Filter capabilities are especially important as concern grows over the adverse health effects of particles from 0.01 to 3.0 micrometers in diameter. Despite their excellent efficiency, fabric filters are often supplanted by more compact, less expensive collectors. If fabric filters could be made more compact, while preserving their superior efficiency characteristics, a considerable savings in first cost might result. Such filters could then be applied where they now are not, to the betterment of air pollution control.

There is no reason why high pressure drop fans such as have been used for the Venturi scrubber cannot be used for a fabric filter as well. The subsequent increase in filtration velocity would reduce the number of filter bags required to process a given quantity of dirty gas. The reduction in filter size and concomitant initial cost would of course have to be balanced by an increase in operating cost due to the larger fan required.

Currently, there is little knowledge of a fabric filter's collection efficiency as a function of its primary operational variables. Even less information is available concerning the mechanisms by which particles collect in or pass through a fabric filter. Although media (fiber) filtration theory is often applied to describe the performance of fabric filters, as more information concerning the operation of fabric filters becomes available the inadequacies of media theory are becoming clear. It is necessary to understand how and why fabric filters perform as they do before an increase in filtration velocity can be expected to succeed.

This project was undertaken to measure the collection efficiency of a commercial fabric filter when operated at conventional or at high velocity, and to determine the mechanisms by which particles collect in or pass through the filter. The mechanism study provides a basis for understanding the trends in efficiency found.

A pulse-jet cleaned filter with two polyester felt bags was tested at face velocities from 5 to 15 cm/s using resuspended flyash as a test dust. Efficiency was measured against face velocity, deposit thickness, and particle diameter by collecting samples on membrane filters for examination under an optical microscope. Efficiency was found to decrease substantially with increasing velocity, to increase with increasing deposit thickness, and to remain relatively constant for all particle sizes from 0.30 to 5.10 micrometers in diameter. Pressure drop increased rapidly with increasing velocity, all else remaining constant. At the highest velocity tested, 15 cm/s, and at the greatest deposit thickness, 70 micrometers, pressure drops exceeded 100 cm of water.

Speculation in the literature concerning the mechanisms governing particle collection in a fabric filter has at times touched on the possibility that deposited dust can make its way completely through the fabric. If this hypothesis is correct, some dust which passes through the filter does so by not remaining collected once deposited, rather than by not becoming collected at all. This distinction is important, and forms the key to understanding the trends in collection efficiency found.

Three mechanisms were postulated by which particles can pass through a fabric filter: straight through penetration, seepage, and pinhole plug generation. In straight through penetration, particles pass through the filter without stopping. Seepage, or bleeding, is a mechanism by which deposited dust gradually works its way through from the dirty to the clean side of the fabric, aided by the drag force exerted on the particle deposits by the gas going past, and perhaps by the deflection of fabric fibers beneath the deposit as the fabric stretches from increasing pressure drop. The pinhole plug mechanism postulates the spontaneous release of a small core of dust from the dust deposit, perhaps at the site where the supporting synthetic felt was needled at the time it was manufactured, although this phenomenon has been noted for woven fabrics as well. Particles can therefore pass through the filter uncollected by the straight through mechanism, or move through the filter from their site of initial deposit by seepage or pinhole plug loss.

The size distribution of particles passing through the filter by the straight through mechanism should reflect a dependence on the forces causing particles to be collected there: inertia, interception, diffusion, gravity, electrostatics, etc. However, the size distribution of the particles which pass through by seepage or pinhole plugs should be the same as the size distribution of the deposited dust, that is, very close to the size distribution of dust fed to the filter.

Using a series of tagged dusts, the proportion of total dust emitted which is accountable to each emission mechanism was measured in relation to face velocity and deposit thickness. Straight through penetration rapidly diminishes in importance, although it is important immediately after cleaning the fabric. The emitted dust accountable to the seepage mechanism is relatively constant during the entire filtration cycle. The pinhole plug mechanism rapidly rises in importance after cleaning, passes through a maximum, and then declines as the dust deposit becomes thicker and pressure drop through the deposit increases. Similar trends were found at each velocity tested.

Trends in efficiency can be interpreted in light of the particle emission mechanisms described above:

(1) A decrease in efficiency with increasing velocity results from increasing dust emissions by the seepage and pinhole plug mechanisms.

(2) An increase in efficiency with increasing deposit thickness results from a decrease in the importance of straight through penetration. This increase more than offsets a simultaneous rise in pinhole plug generation as the dust deposit thickens.

(3) The relative insensitivity of efficiency to particle size results from the predominance of the seepage and pinhole plug emission mechanisms, after straight through penetration has diminished. Seepage and pinhole plug generation should release dust with about the same size distribution as that fed to the filter.

From this work, new insights have been developed into the manner by which fabric filters operate. This knowledge may form the basis for a systematic program focused on improving the performance of conventional fabric filters, and may help to improve the prospects for developing the high velocity fabric filters of the future.

The context in which this thesis was written is described in the section entitled "Background." The main points found in the work carried out are described in the section entitled "Experimental Work." A description of further work needed to bring the work begun here into full fruition is given in a section called "Topics for Further Consideration."

BACKGROUND

Recent appreciation of the hazard from small particles in the atmosphere has caused increased concern over their release from industrial sources (1,2). The several classes of industrial air pollution control equipment used to control particulate emissions have varying abilities to collect these small particles. The devices with lowest initial cost, cyclones and low energy scrubbers (3), depend primarily upon inertial force for particle collection. Unfortunately, inertial force is proportional to the square of particle diameter, and falls off rapidly for the small particles of concern.

Electrostatic precipitators are capable of collecting small particles, but are bulky, and have an initial cost many times that of an inertial collector (3). High energy scrubbers, such as the Venturi, are sometimes able to collect small particles with satisfactory efficiency (4). They are extremely compact, and have a first cost between that of the low efficiency inertial collectors and electrostatic precipitators (3). On the negative side, high energy scrubbers have operating costs an order of magnitude higher than those of most other particle collection devices. The disposal of liquid wastes from these collectors can also be a problem. In addition, some particles, generally those between 0.10 and 0.50 micrometers in aerodynamic diameter and non-wettable in nature, cannot be collected efficiently even when the Venturi employs very high energy inputs.

General agreement exists within the industrial gas cleaning community that filters offer the best collection efficiency of all commercially available collection devices over the entire spectrum of particle sizes delt with in industrial practice, especially for small particles. Cleanable fabric filters make up the vast majority of large scale industrial filter applications. Despite their high efficiency, fabric filters are often not used because of their large size and relatively high initial cost. If fabric filters could be made smaller, both the size and the first cost objections would be reduced.

As conventionally operated, fabric filter bulk is attributable to the low velocity with which dirty gas passes through the filter bags. Filtration velocities are generally maintained to keep the pressure loss across filtration systems at about 12.5 cm of water column, and in line with the pressure losses experienced in most other industrial gas cleaning equipment. There is no apparent reason why filtration velocity cannot be increased through the use of fans similar to those used on high energy scrubbers. As the size of a filter installation is inversely proportional to filtration velocity, a substantial increase in velocity would result in a significant decrease in fabric filter size and first cost, although these advantages would be offset somewhat by an increase in operating cost. A successful reduction in the size and first cost of fabric filtration equipment should allow this high efficiency collector to become competitive with currently used collection equipment which operates with substantially inferior efficiency, especially in the small particle size range.

Perhaps in part because of their excellent efficiency characteristics, fabric filters have not been studied from the theoretical standpoint in the detail which has been practised for other types of air pollution control equipment. As filters have generally been able to achieve even the most stringent emission limitations, there has been less incentive to study how such efficiency was achieved or might be improved. Researchers interested in modeling the collection characteristics of control equipment have turned to other devices. The work done on fabric filters has concentrated on parameters influencing pressure drop rather than efficiency.

Fabric versus Media Filtration Studies

In contrast with the fabric filtration situation, media or fibrous filters have been studied in considerable detail (5-8). Here, the approach has been to identify those methods by which a single particle deposits upon an isolated cylinder. The effects of many particles and fibers in a filter are then combined to give an overall theoretical efficiency for a media filter.

Successful theories describing media filtration are backed by considerable experimental verification. Less effort has gone into studying particle accumulation effects in media filters and, at present, it is not possible to predict from basic theory the relationship between penetration or pressure drop and particle accumulation. Those theories which have been proposed (9) involve the evaluation of an empirical constant from experimental filtration data. Although such theories are useful for developing an understanding of the filtration process, and form a framework for the interpretation of experimental data, they are not useful for predicting the quantitative performance of untried filters on untested dusts.

There are important differences between filtration by a media filter and filtration by an industrial fabric filter. The filtration models which have been used successfully to describe a media filter generally postulate that particles deposit upon clean fiber elements, generally assumed to be cylinders which are distributed randomly throughout the filter. In sharp contrast, within a fabric filter, particles are never presented with clean fibers upon which to collect. Even after cleaning, many particles remain within the fabric structure. After operating for only a short time, enough particles will have deposited upon particles previously collected at the fabric surface to move the primary particle deposition site from the fabric interior to an external deposit or filter cake.

Clean media filters operate at a relatively constant pressure drop, and are often discarded when pressure drop doubles. In contrast, fabric filters operate at a pressure drop that increases rapidly and is returned suddenly to a lower value by periodic cleaning. Directly after cleaning, during the period termed "cake repair" (10) pressure drop rises rapidly while the first new particles deposit on the fabric. As a uniform dust deposit forms on the surface of the cloth, pressure drop increases, although it does so relatively slowly, until a period of "full cake filtration" begins, Thereafter, pressure drop increases at a constant rate with increasing dust deposit thickness.

Despite important differences in the manner by which particles collect in a media as opposed to a fabric filter, it has been repeatedly reported (3,11-18) that the filtration mechanisms which describe media filtration also describe filtration by a fabric filter. Perhaps the mechanisms by which particles collect in and are retained within a media filter also operate in a fabric filter. But once fabric filtration is underway, and pressure drop across the fabric begins to increase, additional mechanisms may come into play in the fabric filter which do not apply in the media filter case. There is currently no available method for predicting the efficiency with which a fabric filter operates (19), nor is there more than conjecture over the means by which particles collect in or pass through a fabric filter (20-21). An evaluation of collection efficiency characteristics and the mechanisms by which particles are emitted by a fabric filter form important parts of this thesis.

Literature Review

Published information is scarce on the collection characteristics of fabric filters. That information which is available must be tempered by evidence (10,22-26) which suggests that changes in cleaning mechanism, fabric type, or dust are likely to change substantially the performance of a fabric filter. Because the importance of each variable has not been well quantified experimentally or theoretically, extrapolations of performance data across various applications must be made with extreme caution. A literature review carried out in conjunction with this project found that no theoretical (20-21) and very little experimental work (27-28) has been carried out to determine the collection efficiency of fabric filters. Of the experimental work that has been reported, only two filters were found comparable to that tested here. This scarcity of reported data should not be construed to imply a lack of concern on the part of workers in this field. Rather, it is indicative of the diversity of fabric filter types and reflects the considerable technical difficulties that must be overcome when measuring fabric filter performance.

Dennis's Studies

Dennis (29) reported experimental data for a pulse-jet fabric filter using polyester felt bags and collecting resuspended flyash. Experimental work was conducted at filter face velocity of 4.3 cm/s, an inlet dust loading of 27 g/m^3, and a filtration cycle of one minute duration. Cleaning was accomplished by a 0.06 second pulse of air at 4.8 atm. Dennis measured inlet particle size distribution using optical microscopy, and converted the count data obtained to a cumulative inlet size distribution expressed on a mass basis. Anderson impactor samples were also taken to determine inlet and outlet mass cumulative size distributions. All size distribution data were evidently obtained from samples taken throughout several filtration cycles. Although confidence intervals for the cumulative size distributions are not given, the three distributions are so similar that they probably are not different statistically. If the inlet and outlet size distributions are in fact the same, collection efficiency for particles of all sizes going to the filter must be constant. Dennis explains his results by postulating that the fraction of fine particles which directly pass through the filter are supplemented by agglomerates loosened during fabric cleaning in just sufficient degree to make the outlet dust size distribution identical with the inlet.

Dennis also reports the results of tests performed on the same filter in which instantaneous particle concentrations were determined at various times within the filtration cycle, for particles of various sizes. A Baush and Lomb light scattering instrument monitored particles greater than 1.0, 2.0, and 3.0 micrometers diameter. The most significant trend observed was that outlet concentration rises sharply immediately after cleaning, and then decreases rapidly during the period of cake repair. After about 0.4 minutes, outlet concentrations leveled off and continued at remarkably constant values throughout the remainder of the filtration cycle. Similar trends were observed for each of the three particle size ranges.

Studies by Billings et al.

Billings, First, Dennis, and Silverman (22) reported the performance of a continuously cleaned fabric filter with wool felt bags when collecting a number of dusts including flyash. The filter was operated at velocities from 5 to 15 cm/s while collecting flyash. Although much of the study was aimed at determining the influence of various parameters on pressure drop and is not of direct concern here, some efficiency measurements were made. Efficiency was not measured as a function of particle size; all efficiency measurements were found by an overall mass procedure. As the reverse-jet cleaning apparatus operated continuously for the flyash tests, no observations of the change in efficiency as a function of time since cleaning could be made. However, tests on atmospheric dust performed without the cleaning mechanism in operation showed that efficiency increased as deposit accumulated on the filter bags. Increased face velocity was found to decrease efficiency, both for atmospheric dust and for flyash.

Inlet flyash concentration was varied over a range from 0.8 to 30 g/m^3. In general, higher inlet concentrations resulted in higher outlet loadings, although the dependence was less than proportional. As inlet loadings became high, outlet loading was found to depend less upon inlet loading.

Untreated wool felt and felts impregnated with silicone or resin were tested to determine the effects of fabric treatment. Both the silicone and resin treated felts gave lower penetration and higher pressure drop than the untreated wool felt. These results were attributed to a reduction in particle migration, or seepage, through the treated fabrics and to enhanced electrostatic attraction between uncollected and collected particles on the treated fabrics.

Other Efficiency Studies

Several other reports of the dependence of efficiency on velocity are available (10,24,30-33). In all cases, for both woven and felted fabrics, efficiency was found to decrease as velocity increased. Williams, Hatch, and Greenburg (34) attribute the efficiency decrease to an increase in pressure drop rather than an increase in velocity, and postulated that if velocity could be increased without influencing pressure drop, no adverse effect on efficiency would be found. Their logic in reaching this conclusion and the method by which the hypothesis could be verified through experiments were not stated.

The effect of deposit thickness on efficiency has not been well quantified. Hall and Cass (35) found that decreasing the time between pulse-jet cleanings of Nomex felt while filtering brass foundry fume resulted in a decrease in efficiency, all else remaining constant. This is consistent with a hypothesis that efficiency increases during the filtration cycle.

Data are scarce regarding the fractional efficiency of fabric filters as a function of particle diameter. McKenna and coworkers (30) showed a minimum in the size versus efficiency curve in the region of 0.30 micrometers diameter for Nomex felt collecting an unspecified dust at 1.5 and

3 cm/s face velocity. They provide no description of how the efficiency data were obtained, no indication of the confidence intervals surrounding their values, nor whether the reported minima are real. Turner (24) showed similar trends for filters treating the effluent from a lead sinter machine, an industrial boiler, and a utility boiler. Turner hedged on the amount of confidence which can be placed in his data, and concluded only that the data "do imply good efficiency in the test (particle size) range."

Peterson and Whitby (26) tested the efficiency of clean and loaded fabrics with monodisperse dye aerosols. They found that cleaned filter efficiency dropped off rapidly for particles less than about 1.0 micrometer in diameter. When the filter was loaded with flyash, its efficiency was relatively constant over the entire particle diameter range tested, from 0.065 to 1.0 micrometers diameter. When the filter was loaded with AC coarse or fine dust, the filter showed minima in collection efficiency in the region of 0.30 micrometers diameter. For Peterson and Whitby's experimental data, efficiency was determined by examining simultaneous concentrations of fluorescent dye aerosol upstream and downstream of the filter. Dust coming from the filter deposit which had been previously laid down on the test fabric could be present on the downstream filter sample, but would not be detected as it would not fluoresce.

Particle Collection and Emission Mechanisms

The mechanisms by which particles pass through a fabric filter have been the subject of much speculation. In general, dust which collects within the filter has been assumed to remain at its site of deposition. Still, in recent years, an appreciation has grown that dust may pass through the filter even after it has been deposited.

Williams, Hatch, and Greenburg (34) stated as early as 1940 that excessive pressure drop across the fabric filter might cause the deposit cake to break down, and lead to the discharge of excessive dust. Billings and coworkers (22) recognised in 1954 the problem of dust deposit seepage in the fabrics they tested, as noted above. "Seepage" was also mentioned by First and Silverman (33) in 1963 as a method by which dust could work its way through a felt fabric. There has been a growing realization that more knowledge is needed on this point, culminating with Caplan's statement (20) in 1974 that "seepage" or "bleeding" must be investigated if fabric filter performance is to be understood.

Stephan, Walsh, and Herrick (36) were the first to study the behavior of a dust deposit during filtration. They found that tiny pinhole punctures developed on the surface of a filter deposit supported by cotton sateen cloth and were easily visible under a microscope. They postulated that such punctures could be caused by static pressure differentials across the deposit, by kinetic energy variations and viscous drag within the cake, or by extraneous physical contacts with the filter. Similar pinholes have been noted by Draemel (37) and Rudnick (38).

Mohamed (39) and coworkers found when testing a needled felt that many small dots appeared on the clean side of the fabric. They interpreted these dots to be the primary sites at which dust passed through the filter, and speculated that the dots might be locations at which the felt was needled at the time of its manufacture.

Purpose of This Study

Although considerable information has been reported on the efficiency characteristics of fabric filters, the data are badly fragmented. No unifying theory has been developed to explain the data, and little reliable information is available to describe the mechanisms by which particles pass through a fabric filter. Therefore, this work was carried out with the following purposes:

(1) to determine the trends in collection efficiency of a pulse-jet fabric filter collecting flyash as functions of velocity, particle diameter and deposit thickness, and

(2) to determine and quantify the mechanisms by which particles pass through the filter.

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EXPERIMENTAL WORK

Introduction

Because of their high efficiency, fabric filters are widely recommended for the control of particulate emissions from diverse industrial processes (1-3). High filter efficiency is especially important for submicrometer particles, which are increasingly suspected to cause severe adverse health effects (4,5). Yet too little is known about how and why fabric filters perform as they do to exploit their potential fully.

Conventional fabric filters are bulky due to the low gas velocity through the filter bags; many bags are required to process a sizeable gas flowrate. One way to decrease the size of a filter installation is to use fewer bags and simultaneously increase the gas velocity through each. Although a more powerful fan must be used to overcome the increased pressure drop generated, such fans have been used often on scrubbers in recent years (6), and there is no apparent reason why they connot be used on fabric filters as well. A pulse-jet filter with polyester felt bags was selected for study, as units of this type are frequently used in industry, yet are able to operate at higher than conventional velocity.

It is important to establish the factors influencing collection efficiency in a control device in order to understand its operation and apply it appropriately. With cyclones, scrubbers, and electrostatic precipitators, these factors have been established fairly well (6-8). They have been found to include particle size, mode of operation, and parameters describing equipment configuration. For fabric filters, these factors have been less well established, as few comprehensive measurements of filter efficiency have been made.

The mechanisms governing particle collection in a fabric filter are often assumed to be identical with those governing collection in a media (fibrous) filter (1,3,9,10), regardless of obvious differences in their construction and operation, and despite the absence of supporting experimental evidence. This work was carried out to establish the efficiency characteristics of a pulse-jet fabric filter when operated at conventional or higher velocity, and to determine the mechanisms by which particles pass through it.

Equipment

A commercial two bag filter was used, identical to industrial equipment in all respects except for its limited number of bags. Figure 1 is a schematic representation of the experimental apparatus. Dust from a vibrating feeder (11) flowed onto a variable speed turntable, where an adjustable baffle formed it into a strip of constant size and swept the excess to waste. After aspirating this dust strip, a pneumatic injector projected it into the inlet air duct. The dusty air passed through a Venturi flowmeter, then flowed past an upstream sampling station, around an inlet baffle, and into the fabric filter housing.

The filters were two polyester, needled felt cylinders, each 11.4 cm in diameter and 240 cm long, made from 16 micrometer diameter fibers of uniform diameter. The fabric weighed 540 g/m^2 when new; each filter had an effective area of 1.72 m^2. The bags were stretched over wire cages to prevent their collapse while filtering from the outside inward. The fabric was conditioned by operating with about 1 g/m^3 dust concentration for several hundred hours before experimental measurements were made. Cleaned air from the filter passed a downstream sampling station, flowed through a fan, flow regulator, and an acoustic muffler, then entered the waste air system.

Electrostatically precipitated flyash was chosen as the test dust for all experiments because of its importance as a pollutant (4,12), its potential for future fabric filter applications (13,14), and its size distribution, which includes a substantial fraction of small particles (15). One batch of flyash was prepared and used for all tests, by sifting it through a 100 micrometer screen to remove extraneous material, and then drying it at 100 °C for several days. The size distribution of the airborne flyash, expressed on a count basis, is given in Figure 2 with 95% confidence intervals (16) from measurements taken upstream of the filter.

Penetration Experiments

Although the experiments and results described here are specific for the fabric, cleaning mode, and dust used, the trends in penetration found may be more properly generalized. For this reason, the interpretation of experimental results presented here is more oriented toward trends in penetration than toward specific numerical values.

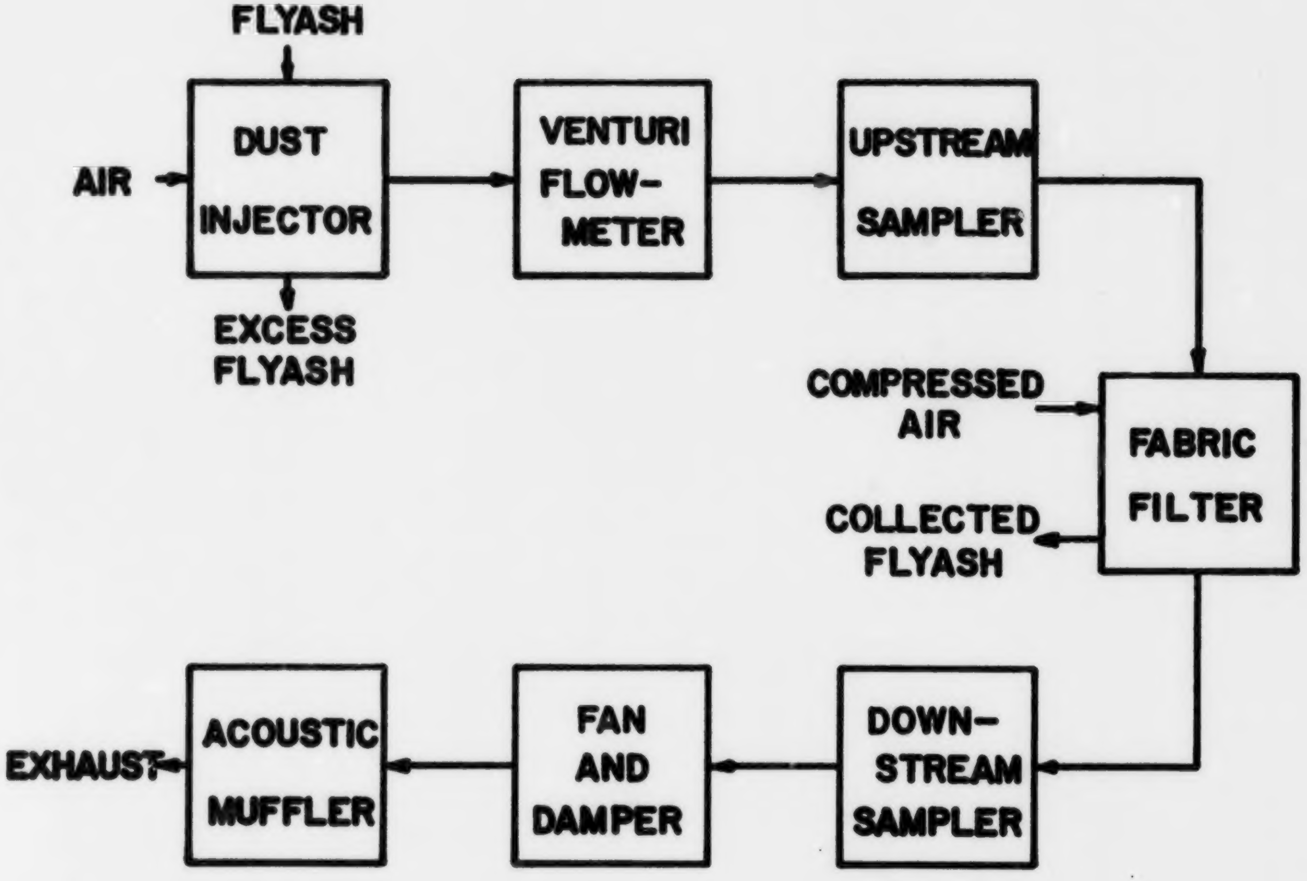
FLYASH
AIR
DUST INJECTOR
EXCESS FLYASH
VENTURI FLOW-METER
UPSTREAM SAMPLER
COMPRESSED AIR
FABRIC FILTER
COLLECTED FLYASH
DOWN-STREAM SAMPLER
FAN AND DAMPER
ACOUSTIC MUFFLER
EXHAUST

Figure 1 Schematic of Experimental Fabric Filter Apparatus

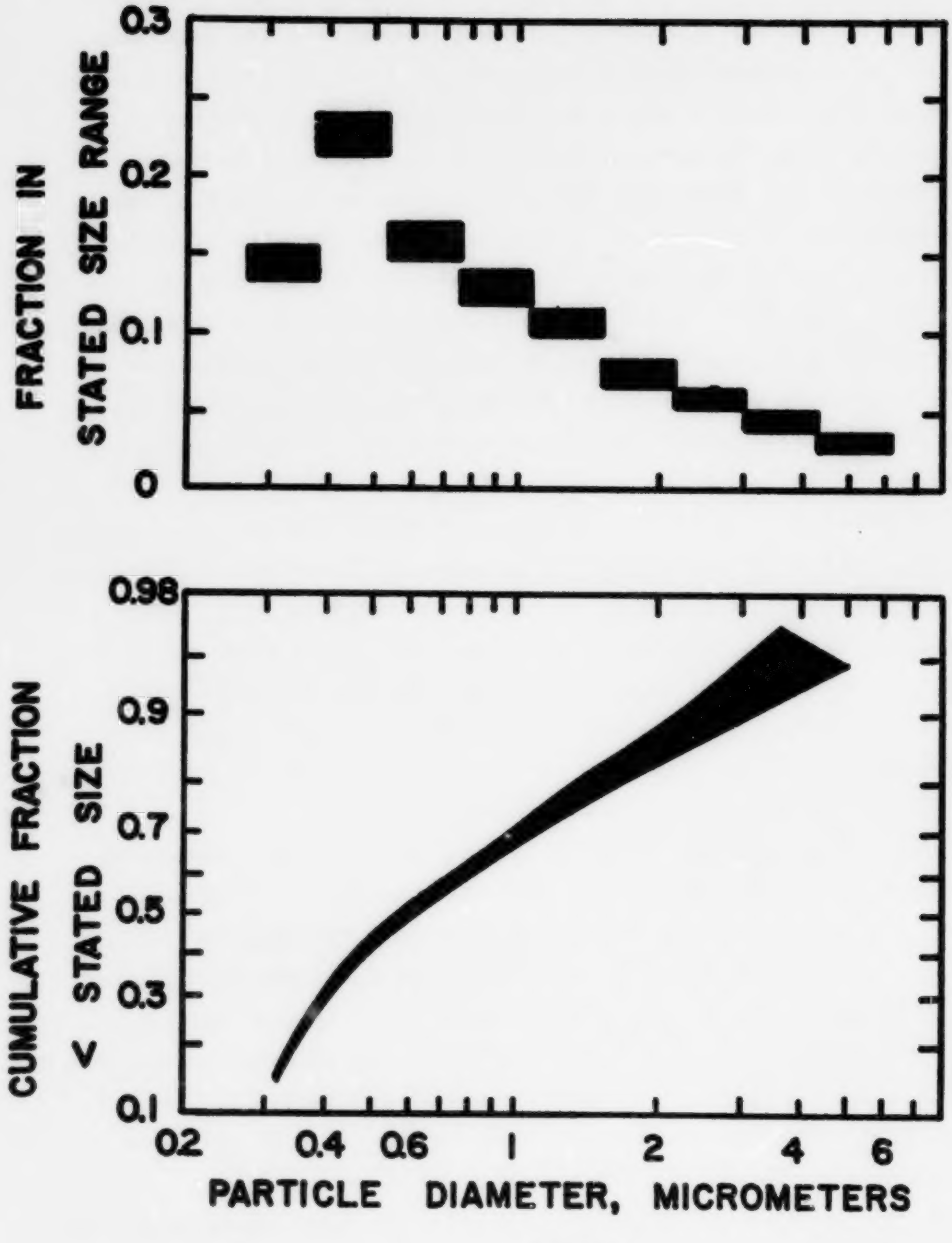

Figure 2
Fractional (above) and Cumulative Flyash Particle Size
Distributions, with 95% Confidence Intervals

Experimental Measurement of Penetration Penetration was measured by taking isokinetic samples on membrane filters upstream and downstream of the fabric collector. Microscope slides prepared from the membrane filters (17) were examined using a 90 X oil immersion objective and a 25 X eyepiece fitted with a Porton-May counting and sizing graticule, to determine the size distribution of upstream and downstream dust. Penetrations were then calculated for each of nine particle sizes, ranging from 0.27 to 4.32 micrometers in diameter. The diameter of each particle was defined as the diameter of a sphere having the same projected area (18). From 500 to 1000 particles were sized on each slide, and 95% confidence intervals were determined for all penetration measurements (19). All data were corrected for stray particles on the slides or in the mounting solutions. Additional data on the flyash used and on the operation of the filter are given in Table 1.

Face velocities from 5 to 15 cm/s through the filter bags were tested. Figure 3 is a plot of pressure drop against deposit thickness with velocity as parameter for these tests. Operation was impossible at velocities in excess of 15 cm/s as pressure drops exceeded 100 cm of water. During all tests, a dust sample was taken during each two minute time interval from zero to 16 minutes after cleaning. All together, 360 values of penetration were found, as functions of particle diameter, velocity, and time since cleaning, i.e., deposit thickness.

Cleaning followed accepted industrial procedures. After completing a test, pulses of air at 6.8 atm were introduced at the mouth of each filter bag; no cleaning took place during sampling. The shock of the air pulse coupled with possible reverse air flow through the fabric achieved cleaning. The inlet dust concentration was kept at 0.60 g/m^3 for all tests. Normally (23), commercial equipment operates at dust concentrations ten times or more higher than allowed in these experiments; the time between cleanings is usually about one tenth as long as that used here. Low dust concentration and extended time between cleanings allowed enough expansion of the experimental time scale to make the experiments manageable.

Inlet dust concentration, c_e, face velocity, v, time since cleaning, t, and flyash bulk density, ρ , can be related to theoretical flyash deposit thickness, X, through Equation 1 after making the reasonable

Table 1

Operational Data for Fabric Filter Experiments

Sampling times : 2, 4, 6, 8, 10, 12, 14, 16 minutes

Particle diameters : 0.27, 0.38, 0.54, 0.76, 1.08, 1.52, 2.16, 3.04, 4.32 μm

Velocities : 5, 7.5, 10, 12.5, 15 cm/s

Flyash particle density : 2.2 g/cm^3

Flyash bulk density : 1.2 g/cm^3

Inlet dust conc'n : 0.60 g/m^3

Fabric type : Polyester needled felt

Fabric fiber diameter : 16 μm

Fabric weight : 540 g/m^2

Bag size : 11.4 cm diameter, 240 cm long, each of two

Bag area (effective) : 0.86 m^2, each of two

Pulse jet pressure : 6.8 atm

Cleaning interval : 16 minutes

D. B. temp : 18° C

W. B. temp : 10° C

% Relative humidity : 32

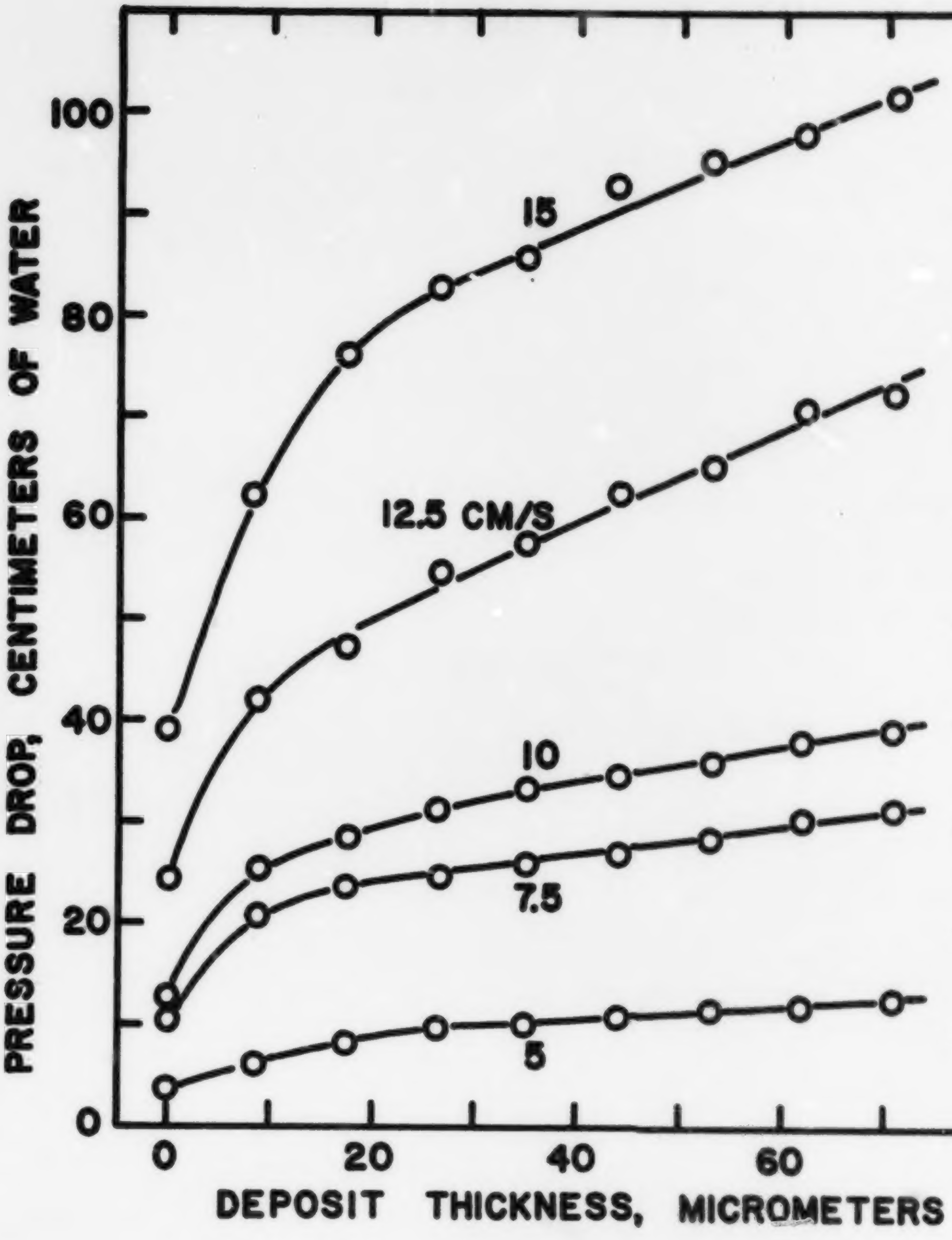

Figure 3
Pressure Drop versus Deposit Thickness, Velocity as Parameter

assumption that the fabric filter operates at a fractional efficiency near unity.

$$X = \frac{c_e \, v \, t}{\rho} \qquad 1$$

Other investigations (20-22) have revealed wide variations in the dust deposit mass per unit area from point to point on woven filter bags, and similar variations may occur on the felt bags used here. The flyash bulk density measured prior to aerosolization, as reported in Table 1 and used for finding deposit thickness, may not be the bulk density of the flyash after deposition on the fabric. Therefore, the theoretical deposit thickness, X, calculated from Equation 1 is at best only an index of the actual deposit thickness at any point.

Results of Penetration Measurements An analysis of variance identified trends in the 360 penetration values. The following trends are statistically significant at the 99% confidence level:

(1) Penetration decreases with increasing time since cleaning.
(2) Penetration increases with increasing log of particle diameter.
(3) Penetration increases with increasing face velocity.

Figure 4 is a plot of penetration against time since cleaning with 95% confidence intervals indicated. Each of the eight penetration values plotted is the result of 45 measurements over the complete range of particle sizes and filtration velocities investigated. Figure 5 shows penetration against particle diameter, and Figure 6, penetration against filtration velocity.

The initial sharp decrease in penetration with increasing filtration time (deposit thickness) was expected from field experience and laboratory data (1,9,23-24). As deposit thickness increased further, penetration leveled off. Similar trends in penetration with deposit thickness have been reported for a pulse-jet filter also collecting flyash on a polyester felt bag (25), and for shake-cleaned filters collecting flyash on woven bags (24,26).

Experiments with industrial filters using woven fabrics indicate that penetration passes through a maximum as particle diameter increases

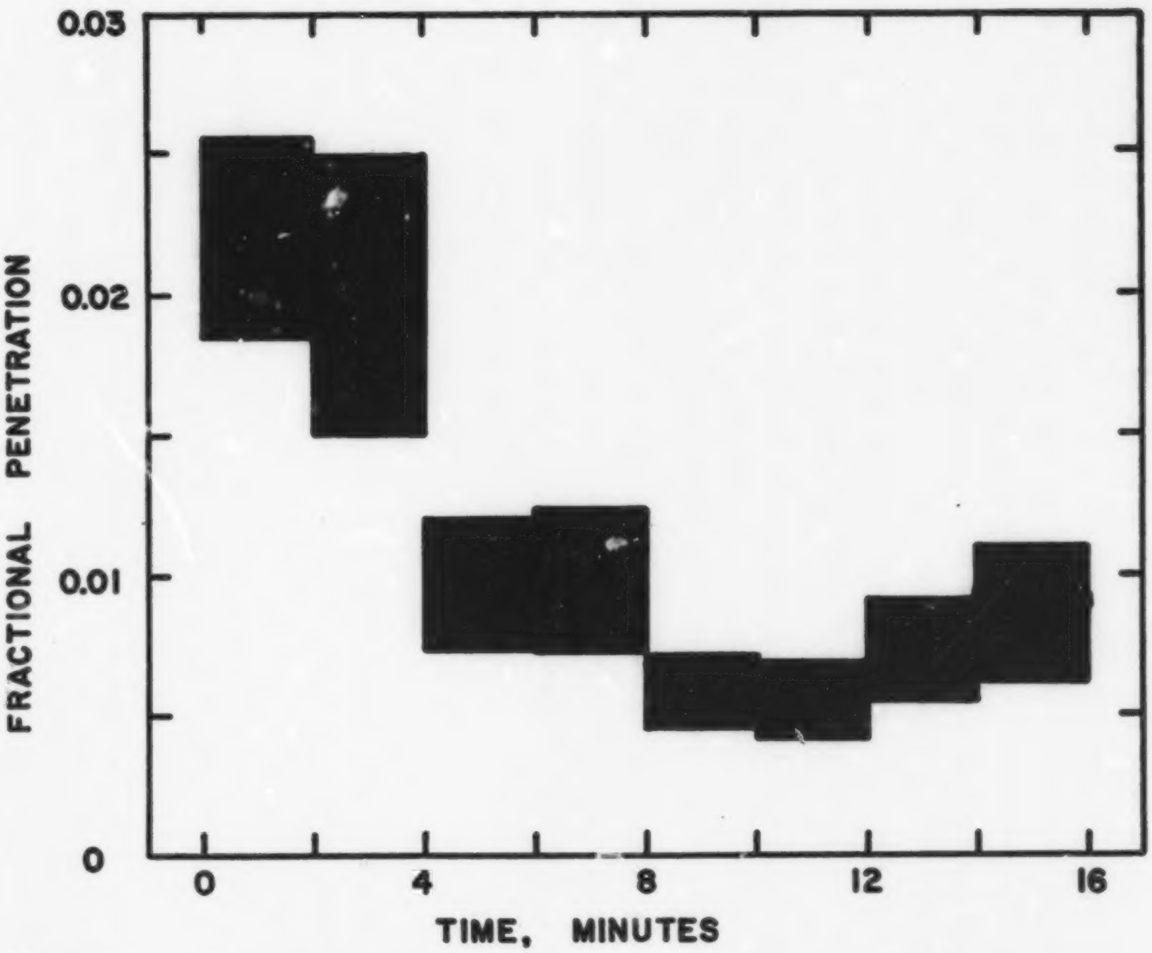

Figure 4 Fractional Penetration versus Time, with 95% Confidence Intervals

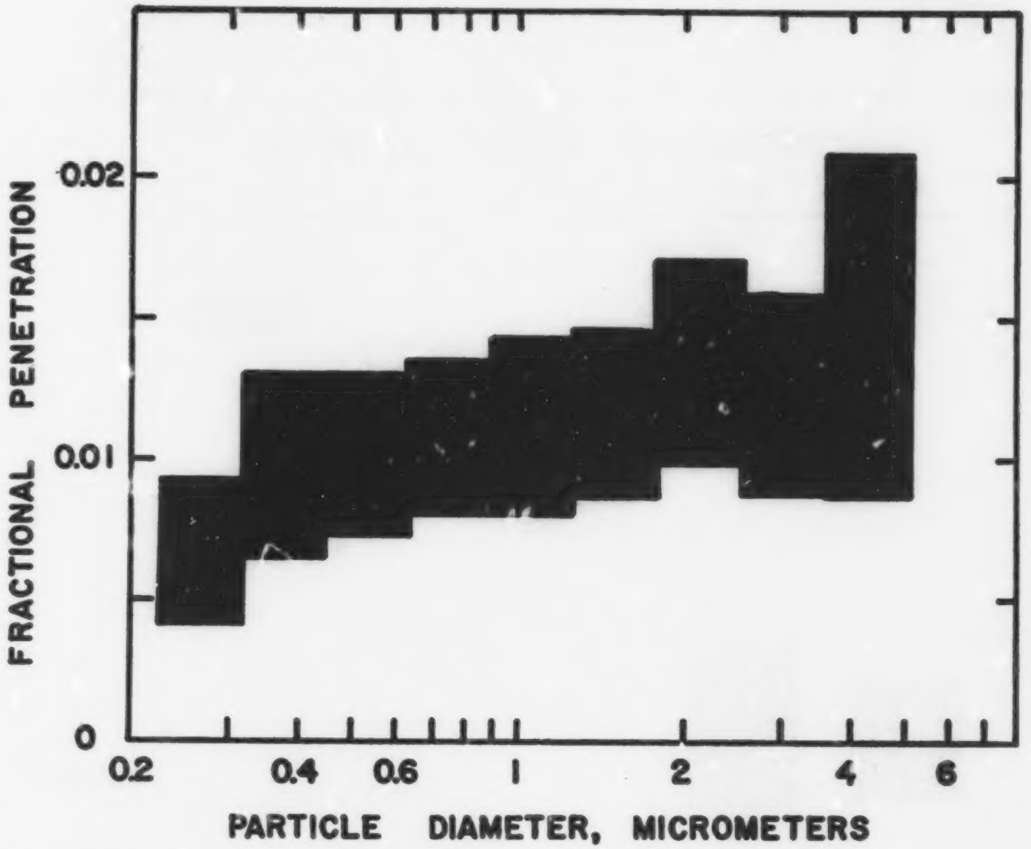

Figure 5 Fractional Penetration versus Particle Diameter, with 95% Confidence Intervals

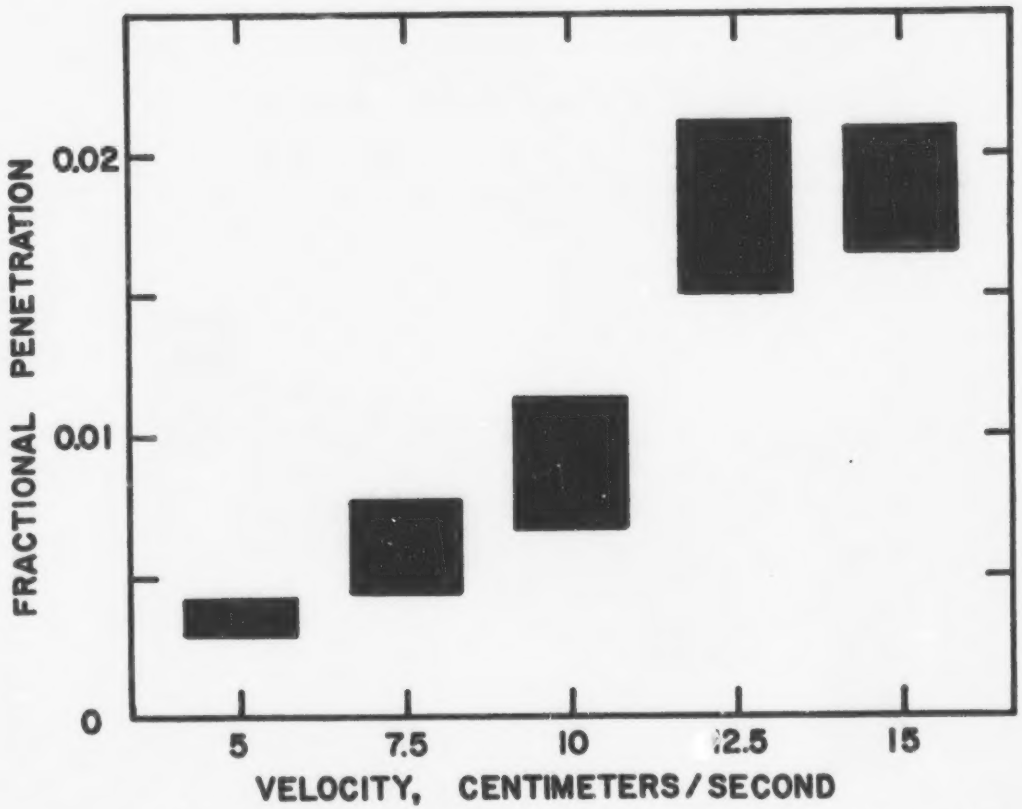

Figure 6 Fractional Penetration versus Face Velocity, with 95% Confidence Intervals

from 0.1 to 10 micrometers (13, 27-28). Although fewer results have been reported for needled felt filters, one study has found a similar size distribution into and out of a pulse-jet cleaned unit, suggesting equal penetration over the 0.5 to 40 micrometer particle diameter range studied (25). Figure 5, the plot of penetration versus particle diameter, shows that penetration increased only slightly with increasing particle diameter. Overall penetration, the weighted average for all particle sizes, is therefore a good indicator of fabric filter collection capabilities.

Other laboratory data for both woven and felt filters confirm the trend shown in Figure 6, that penetration increases with increasing face velocity (13,26-27,29-32). Although previous investigators have not explored this dependence in depth, the data presented here suggest that the increase in penetration with velocity is substantial.

Media filtration theory has often been used to describe the collection of particles in a fabric filter (1,3,9-10). Inertial impaction, interception, diffusion, gravity, and electrical effects have been studied in detail (33-36) and have been shown to be the means by which particles collect upon clean fibers. In media theory, particles not collected by one or another of these effects are assumed to pass entirely through the filter uncollected, whereas collected particles are assumed to remain collected.

The trends in penetration shown in Figures 3-6 disagree with trends predicted by media filtration theory. Media, or clean fiber theory, cannot predict from fundamentals the effect of increasing dust deposit on penetration, as shown in Figure 4. Media theory cannot account for the relative insensitivity of penetration to particle size shown in Figure 5. An increase in penetration with increasing velocity as shown in Figure 6 could be explained by single fiber theory if diffusion were the dominant collection mechanism. However, diffusion should be unimportant for particles larger than a few tenths micrometer in diameter, and most of the flyash particles tested here are much larger than this.

Media filtration theory has been used to predict the efficiency of unused felt fabrics for extremely low inlet dust concentrations (37). However, the collection situation in fabric filters as conventionally used is far different from the ideal, clean fiber situation postulated for the proper application of media theory. In conventional fabric filtration, most dust collects on previously deposited particles rather than on clean cylindrical fibers. Because media filtration theory does not describe the trends in penetration found in fabric filters, and was not developed for the operational conditions found there, it should not be used to predict or interpret the penetration characteristics of fabric filters.

Emission Mechanism Experiments

Immediately after cleaning, many particles collect upon the exposed felt fibers. Soon, however, a continuous dust deposit forms on the fabric surface and particles collect upon previously deposited dust (9,38-39). Particles not collected by the filter, but which pass through without stopping will be said to penetrate the filter by the "straight through" mechanism.

Once a particle lands on or in the fabric, it need not necessarily remain at its point of initial impact. As the dust deposit builds up, pressure drop can increase to several times its initial value as shown in Figure 3, the plot of pressure drop against deposit thickness. Meanwhile, the fabric substrate may stretch, allowing some previously collected particles to work through (40). Filter behavior of this sort will be called "seepage."

Small diameter pinholes have been found at the surface of a dust deposit on woven fabrics (21,41-42). Similar holes on a needle punched felt such as the one tested here may correspond to the places where needles penetrated the cloth during its manufacture (37). A plug of deposited particles may dislodge from the dust deposit and pass through the fabric all at once as the supporting fibers move and stretch beneath it, leaving behind such a pinhole. Particles which pass through the filter in this way do so by a "pinhole plug" mechanism.

Therefore, particles can pass through the filter by the "straight through" mechanism, without being stopped, whereas previously collected particles can make their way through by the "seepage" and "pinhole plug" mechanisms. Figure 7 is a schematic representation of particle emission by each of these means.

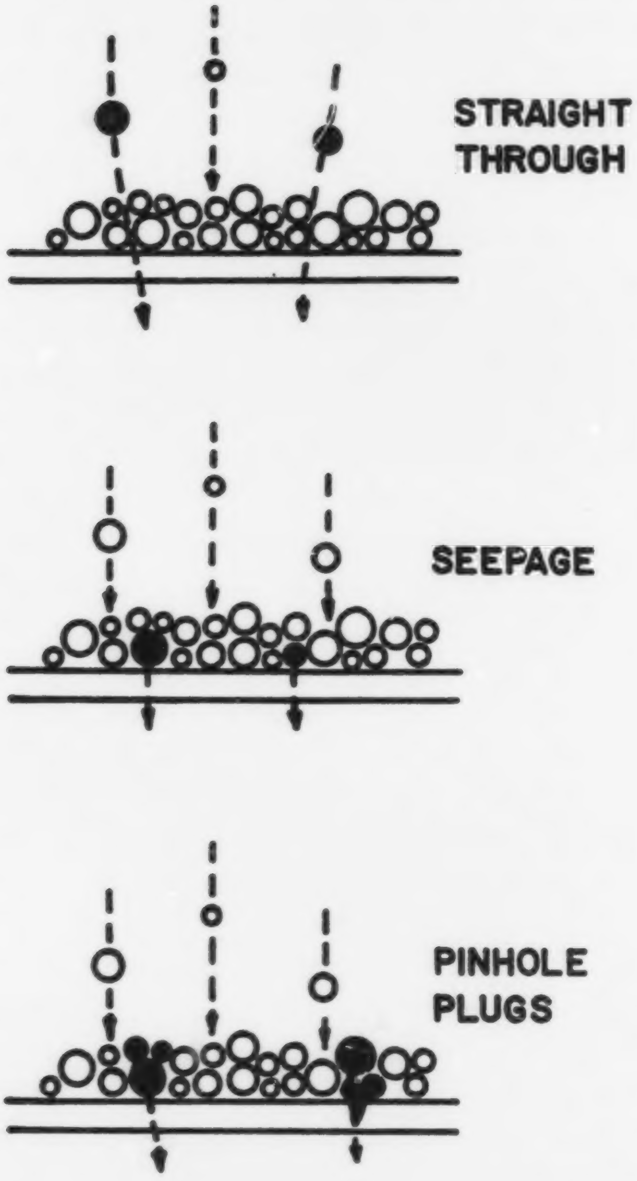
STRAIGHT THROUGH
SEEPAGE
PINHOLE PLUGS

Figure 7

Schematic Representation of Flyash Emission Mechanisms

Experimental Measurement of Emission Mechanisms To determine the importance of these three emission mechanisms, three types of chemically tagged flyash were aerosolized and fed to the filter in the sequence given in Table 2. Operation of the fabric filter for these experiments was the same as for the penetration experiments previously described. The flyash was tagged by mixing it with enough magnesium, lithium, or manganese sulfate dissolved in water to make dust containing 2% by weight of the metal ion after evaporation *.

Downstream dust samples were taken on fiberglass filter paper rather than on membrane paper to minimize pressure loss across the filter paper and permit sampling at higher flowrates. Before use, the clean fiberglass paper was extracted with 0.05 M HCl to remove all residual quantities of the metal tracer ions. After sampling, the amount of each metal tracer present in the downstream air was determined by placing the filter paper with dust deposit in 0.05 M HCl to free the tracer ions, and analyzing the solution by atomic absorption flame photometry. This technique can detect quantities of tagged flyash as small as 0.03 mg, and is specific for each metal ion. Magnesium, manganese, and lithium were selected from many potential chemical tags through the consideration of analytical sensitivity, tagged dust physical properties, tag toxicity, and potential for analytical interference problems.

The proportion of each tag in the total flyash downstream of the filter will depend upon which of the emission mechanisms are operating. If dust penetrates by passing straight through the filter, it should carry the same tag as the dust fed to the filter at that time. If dust seeps through, it should be tagged the same as the dust initially fed. If dust is emitted by the pinhole plug mechanism, it should be tagged in the same tag proportion as the dust fed up until the time when the plug breaks away and slips through.

Table 3 lists the relative proportions of each metal tag expected in the downstream dust according to each dust emission model, at various times after cleaning. Dust samples taken during each of the two minute intervals listed in Table 3 were analysed to find the amount of each of the three tracer metals present. The total dust emission was found by adding the amounts of each tagged dust present, so that the proportion

* See note, page 54

Table 2

Sequence in which Tagged Dusts were Fed

Time	Tagged Flyash Injected
0 - 1 minutes	magnesium
1 - 1½	manganese
1½ - 4	lithium
4 - 8	manganese
8 - 14	lithium
14 - 16	manganese

Table 3

Emitted Dust Tags Expected

Time, Minutes	Tag	Fraction of All Dust Tagged Straight Through Model	Seepage Model	Pinhole Plug Model
0 - 2	Mg	0.50	1.0	0.847
	Mn	0.25	0	0.119
	Li	0.25	0	0.034
2 - 4	Mg	0	1.0	0.347
	Mn	0	0	0.173
	Li	1.0	0	0.480
4 - 6	Mg	0	1.0	0.203
	Mn	1.0	0	0.290
	Li	0	0	0.507
6 - 8	Mg	0	1.0	0.144
	Mn	1.0	0	0.500
	Li	0	0	0.359
8 - 10	Mg	0	1.0	0.112
	Mn	0	0	0.502
	Li	1.0	0	0.386
10 - 12	Mg	0	1.0	0.091
	Mn	0	0	0.410
	Li	1.0	0	0.499
12 - 14	Mg	0	1.0	0.077
	Mn	0	0	0.347
	Li	1.0	0	0.576
14 - 16	Mg	0	1.0	0.067
	Mn	1.0	0	0.366
	Li	0	0	0.568

Coefficients listed for the pinhole pulg model were obtained by averaging the tag makeup of the plugs over the two minute sampling interval indicated, using the mean value theorem.

of this total represented by each tag could be found. The fraction of total emitted dust accountable to each emission mechanism was calculated by solving the appropriate set of three mass balance equations, one for each tag, and three unknowns, one for each emission mechanism. An example calculation is shown in Table 4 where dust collected from 10 to 12 minutes was found to be tagged 24% with Mg, 24% with Mn, and 52% with Li. Using the coefficients from Table 3 for this time period, 23% of the emitted dust was found to be accountable to the straight through mechanism, 19% to seepage, and 57% to pinhole plug generation.

Results of Emission Mechanism Measurements An analysis of variance found significant trends in the dust emission mechanism data. The time-interaction was highly significant. As time increases and the dust deposit thickens, the mechanisms change by which dust is emitted. The velocity-mechanism interaction was not significant. At any fixed time, the fraction of dust emitted by each mechanism is fairly constant from velocity to velocity.

To quantify the proportion of the total flyash emitted which is accountable to each emission mechanism, three polynomial regression equations were fitted to the mechanism data (43-44). These equations predict the fraction, f, of the total dust emitted which is accountable to each of the three emission mechanisms at any dust deposit thickness, X. No velocity terms are included in the regressions as the effect of velocity is not significant. These regressions included the constraints that all predicted values of f must be greater than zero and less than unity, and that the sum of the three f values at any one deposit thickness must be unity. In equations 2-4, deposit thickness, X, is expressed in centimeters.

$$f_{st} = 8.57 \times 10^{-1} - 8.37 \times 10^{2} X + 2.51 \times 10^{5} X^{2} - 2.13 \times 10^{7} X^{3} \qquad 2$$

$$f_{se} = -1.77 \times 10^{-2} + 2.23 \times 10^{2} X - 4.16 \times 10^{4} X^{2} + 1.72 \times 10^{6} X^{3} \qquad 3$$

$$f_{pi} = 1.60 \times 10^{-1} + 6.14 \times 10^{2} X - 2.09 \times 10^{5} X^{2} + 1.96 \times 10^{7} X^{3} \qquad 4$$

Figure 8 is a graphical representation of Equations 2-4.

The range of deposit thicknesses for which the regressions were made is from about 2 micrometers to about 60 micrometers, and extrapolation

Table 4

Example Calculation for Emission Model Determination

			Emission Model				
Fraction of Dust Tagged As:			Straight Through		Seepage		Pinhole Plugs
Mg:	0.24	=	$0\ X_1$	+	$1.0\ X_2$	+	$0.091\ X_3$
Mn:	0.24	=	$0\ X_1$	+	$0\ X_2$	+	$0.410\ X_3$
Li:	0.52	=	$1.0\ X_1$	+	$0\ X_2$	+	$0.499\ X_3$

$X_1 = 0.23$ = proportion of emitted dust accountable to straight through penetration

$X_2 = 0.19$ = proportion of emitted dust accountable to seepage

$X_3 = 0.57$ = proportion of emitted dust accountable to pinhole plug generation

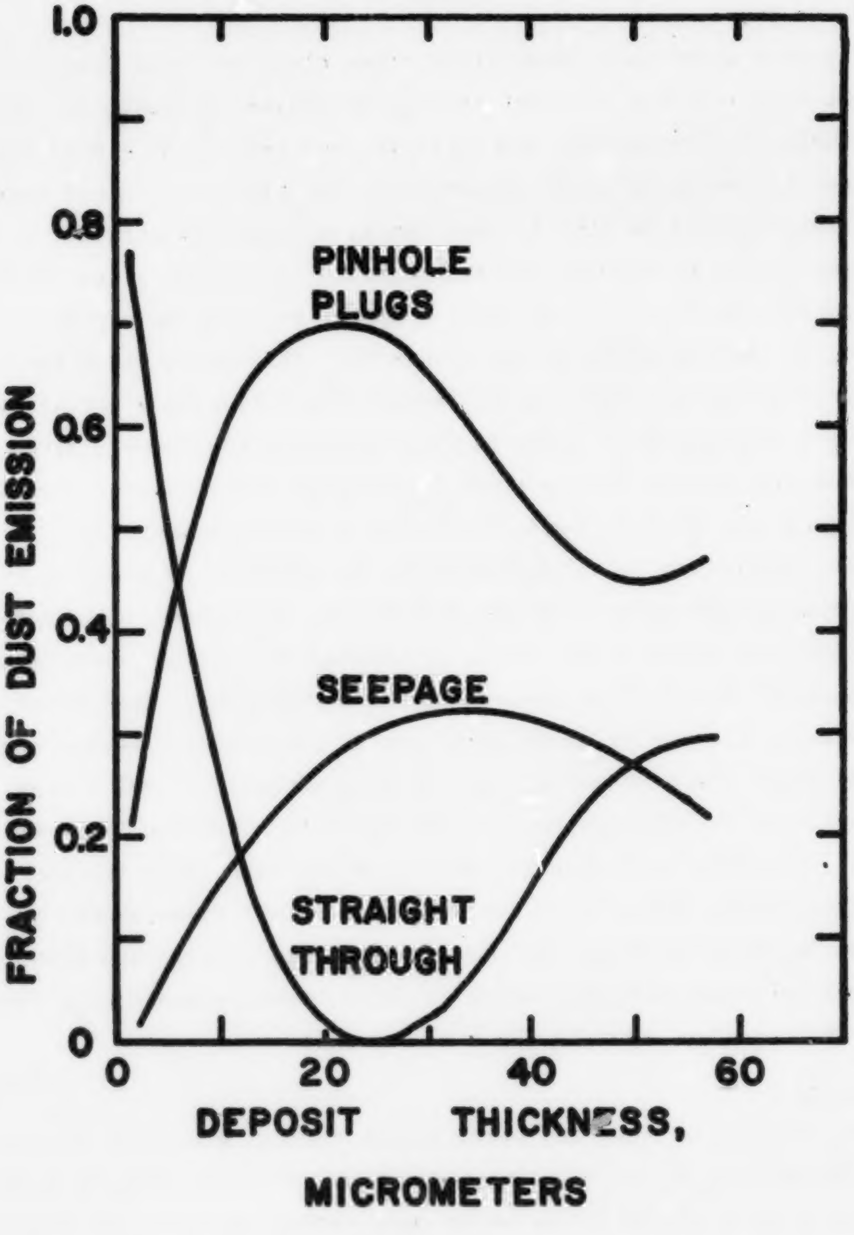

Figure 8

Fraction of Total Flyash Emitted, which is Accountable to Indicated Emission Mechanism, versus Deposit Thickness of the equations beyond this region is not recommended. The correlations between the experimentally determined values of f and those predicted by the above equations are all statistically significant at the 95% confidence level.

Figure 8 shows that immediately after cleaning, most dust comes from the filter by the straight through mechanism, although the seepage and pinhole plug mechanisms soon gain in importance. At a dust deposit thickness of twenty to forty micrometers, the pinhole plug and seepage mechanisms begin to decline in importance, whereas the straight through mechanism begins to become important again. As pinhole plugs leave the dust deposit, the craters left behind appear to open the way for additional emission by the straight through mechanism. This may explain the later importance of the straight through mechanism, following a period of relative dormancy during which the pinhole plug mechanism dominates. Although the major trends shown by Figure 8 are statistically significant, comparatively small variations in the trends are probably not important.

The penetration data trends can now be examined in light of the emission mechanism data. The sharp decrease in penetration immediately after cleaning, shown in Figure 4, is contemporary with a decline in the importance of the straight through emission mechanism. Thereafter, the seepage and pinhole plug mechanisms gain in importance, causing penetration to level off. The seepage and pinhole plug mechanisms should pass deposited particles of all sizes with equal facility. Because these mechanisms dominate emissions from shortly after cleaning through to the end of the filtration cycle, penetration for particles of all sizes should be roughly equal as is shown in Figure 5. Emissions due to seepage and pinhole plugs may increase with increasing velocity, thereby explaining the increase in penetration with velocity shown in Figure 6.

Conclusions

The fraction of dust fed which passes through a control device uncollected is defined as fractional penetration. In most collectors other than fabric filters, particles which pass through do so by not collecting during the brief time they spend there. For this reason, outlet dust concentration is proportional to inlet concentration, as penetration remains constant. However, in a fabric filter, some dust which was initially collected passes through later by the seepage or pinhole plug mechanisms. The primary means of emission then becomes the delayed release of collected particles rather than the passage of particles straight through the filter.

The period of time involved in such delayed release greatly exceeds the time a particle spends in passing straight through without stopping. Therefore, after the straight through mechanism has declined in importance, the outlet dust concentration from a fabric filter should bear little relation to the inlet concentration, as has been previously reported (25-26). Outlet mass flux, the mass emerging per unit filter area per unit time (g/cm^2 s) will be a better indicator of fabric filter behavior than is penetration with its implied proportionality to inlet concentration.

Outlet mass flux, N_o, is related to outlet dust concentration, c_o, and velocity, v, by Equation 5.

$$N_o = c_o v \tag{5}$$

The outlet mass flux attributable to any one of the straight through, seepage, or pinhole plug mechanisms, N_i, can be calculated by multiplying total outlet mass flux by the proportion, f_i, of total emissions attributable to that mechanism from Equations 2-4.

$$N_i = N_o f_i = c_o v f_i \tag{6}$$

The mass flux coming from the filter by each mechanism was calculated using Equation 6, under all conditions for which outlet mass concentrations had been determined during the penetration experiments described above. These data are used for the evaluation of constants in the theoretical equations for mass flux due to each emission mechanism in the following way.

Mass flux due to the straight through emission mechanism, N_{st}, should decrease as the dust layer builds up, for as this layer increases, a particle passing straight through would have to penetrate by a longer and more tortuous path. The relationship between outlet mass flux due to the straight through mechanism, N_{st}, and deposit thickness, X, is described[45] by Equation 7.

$$N_{st} = N_e \exp(-(K_1 X)^{K_2}) \qquad 7$$

Here, N_e is the mass flux entering the filter, while K_1 and K_2 are constants which may depend upon flyash, collector, operational, or other variables.

The mass flux due to the seepage mechanism should depend upon the drag force exerted by the gas flowing past deposited particles. This force should increase with increasing velocity, but not with deposit thickness. At constant velocity, seepage mass flux should be constant at all thicknesses.

$$N_{se} = K_3 \qquad 8$$

Mass flux from pinhole plug generation will depend upon both the thickness of the dust deposit and the probability of plug breakoff. Plug thickness should be proportional to plug mass, and therefore also to outlet mass flux accountable to this mechanism.

$$N_{pi} \propto X \qquad 9$$

However, as the deposit becomes thicker, the additional interparticle bonds which form should make the generation of a pinhole plug increasingly difficult, although increasing pressure drop could act against this trend. This relationship might be described by Equation 10.

$$N_{pi} \propto \exp(-K_5 X) \qquad 10$$

The combination of Equations 9 and 10 gives an expression for the mass flux due to pinhole plug generation, where K_4 is a proportionality constant.

$$N_{pi} = K_4 X \exp(-K_5 X) \qquad 11$$

The total mass flux from the fabric filter, N_o, can be found by adding the contributions of the straight through, seepage, and pinhole emission fluxes.

$$N_o = N_{st} + N_{se} + N_{pi} =$$
$$N_e \exp(-(K_1 X)^{K_2}) + K_3 + K_4 X \exp(-K_5 X) \qquad 12$$

The experimental data for outlet mass flux accountable to each emission mechanism developed from Equation 6 were used with a least squares technique to evaluate constants K_1 - K_5 in the theoretical mass flux equations. Separate evaluations for the constants were made at each velocity and the values found are listed in Table 5. Correlations between data and theoretical equations ranged from 0.86 for the seepage mechanism to 0.91 for the straight through mechanism. As these correlations are high, the postulated forms of the mass flux equations for each mechanism may be suitable, and the assumptions made in establishing their form may be realistic. Figures 9 through 12 are plots of outlet mass flux against deposit thickness at 10 cm/s face velocity according to the straight through, seepage, pinhole plug, and total emission mechanisms.

In an industrial fabric filter, one may be interested in the total mass of dust released to the atmosphere over many filtration cycles, as well as the proportion of small particles in that release. By finding the fraction of total mass released which is accountable to each emission mechanism, the relative importance of each mechanism can be evaluated.

The fraction, z, of total mass released due to any single mechanism can be found from Equation 13 for filtration cycles with cleaning at dust deposit thickness X.

$$z = \frac{\int_0^X N_i \, dX}{\int_0^X N_o \, dX} \qquad 13$$

The mass flux due to any one mechanism, N_i, and the total mass flux, N_o, depend upon deposit thickness, X, and velocity as described by Equations 7 through 12, and the entries in Table 5. Equation 13 was numerically integrated to find the fraction of total mass emitted by each mechanism as a function of deposit thickness at cleaning. Figures 13 and 14 show the results of the integrations for velocities of 5 and 15 cm/s.

At low velocity, as shown by Figure 13, the proportion of all dust emitted which is accountable to the straight through mechanism is large compared to the proportions accountable to the other mechanisms, no matter when cleaning takes place. However, at higher velocity, as shown by Figure 14, the proportion of all dust emitted which is accountable to the

Table 5

Values of Constants for Mass Flux Equations

Velocity cm/s	K_1 1/cm	K_2 dimen'less	K_3 $g/(cm^2\ s)$	K_4 $g/(cm^3\ s)$	K_5 1/cm
5	1.64×10^5	4.09×10^{-1}	1.64×10^{-9}	3.99×10^{-5}	1.43×10^3
7.5	1.12×10^5	3.96×10^{-1}	5.43×10^{-9}	1.23×10^{-4}	1.27×10^3
10	1.66×10^5	3.29×10^{-1}	1.42×10^{-8}	2.90×10^{-4}	1.10×10^3
12.5	4.14×10^8	1.27×10^{-1}	4.06×10^{-8}	2.69×10^{-4}	6.01×10^2
15	1.15×10^6	2.24×10^{-1}	5.97×10^{-8}	3.42×10^{-4}	4.62×10^2

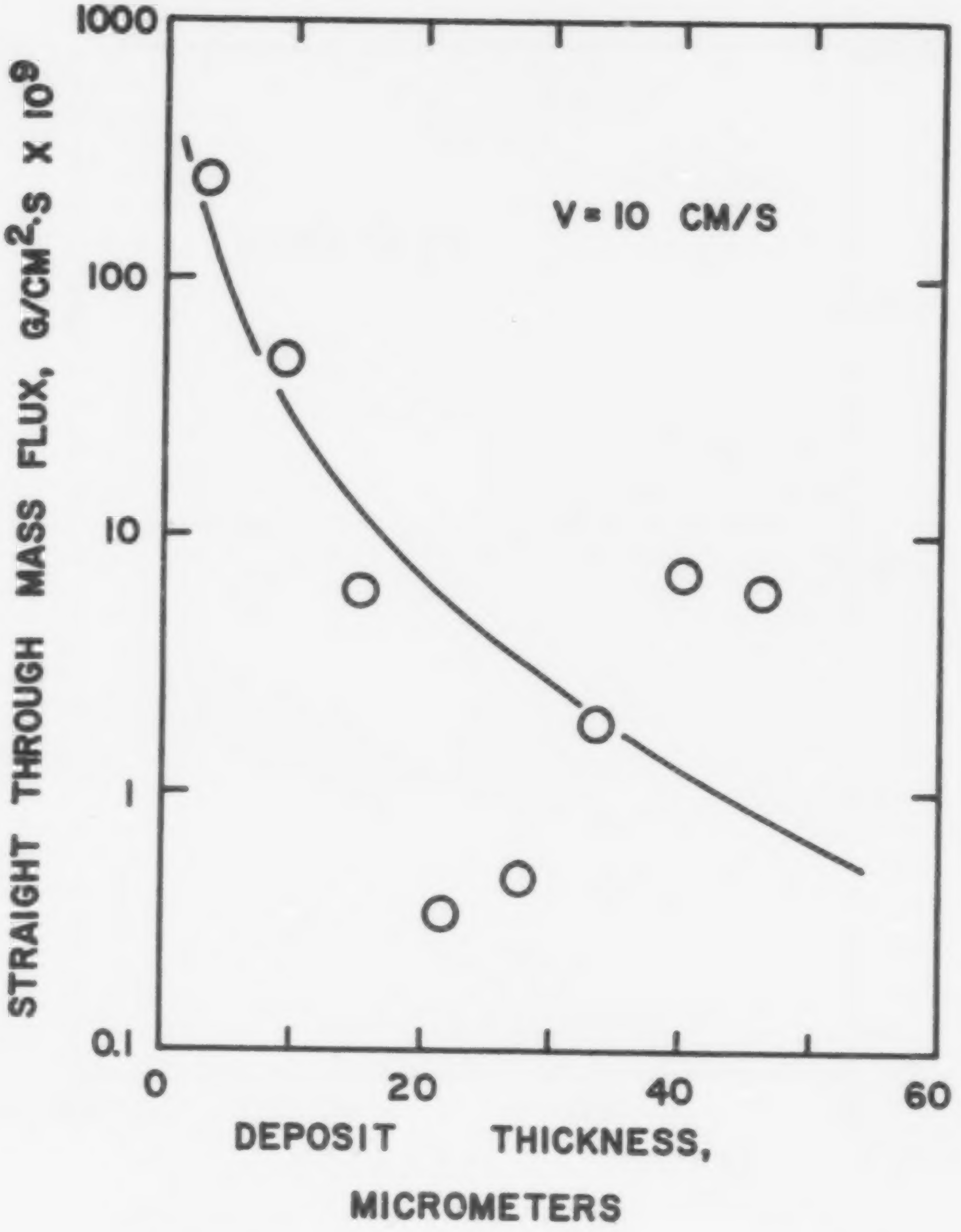

Figure 9

Actual versus Theoretical Mass Flux due to Straight Through Mechanism, Velocity of 10 cm/s

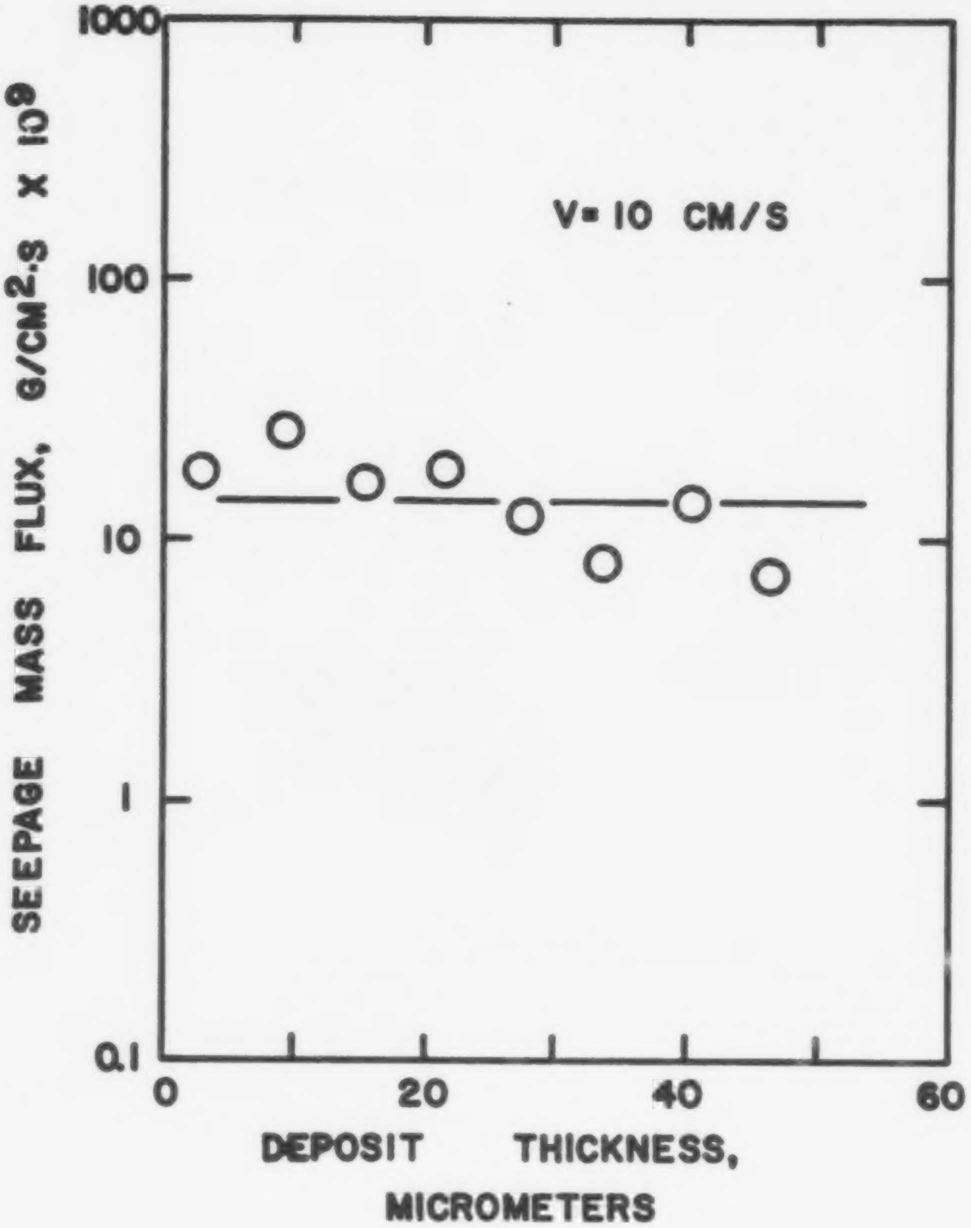

Figure 10

Actual versus Theoretical Mass Flux due to Seepage

Mechanism, Velocity of 10 cm/s

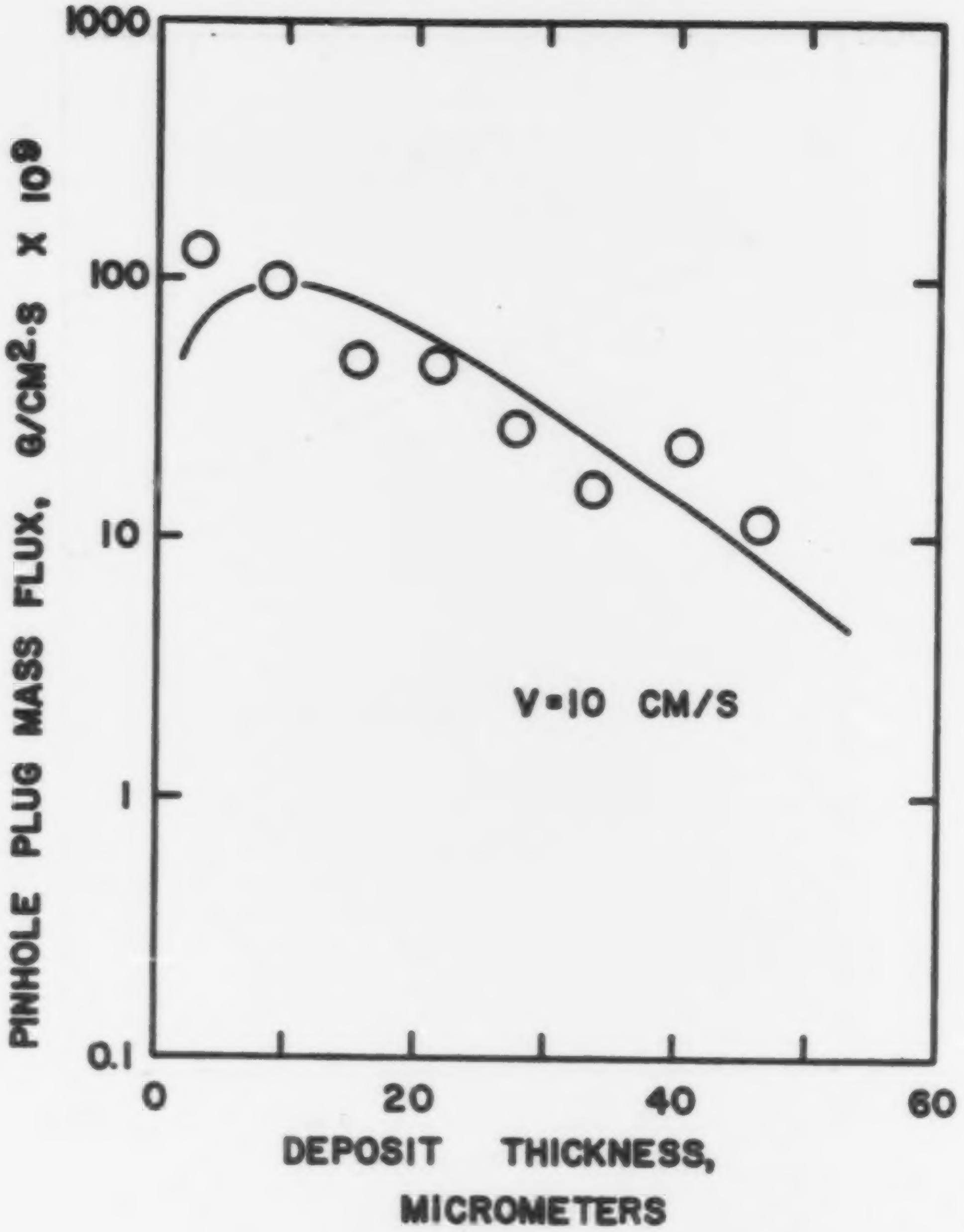

Figure 11

Actual versus Theoretical Mass Flux due to Pinhole Plug Mechanism, Velocity of 10 cm/s

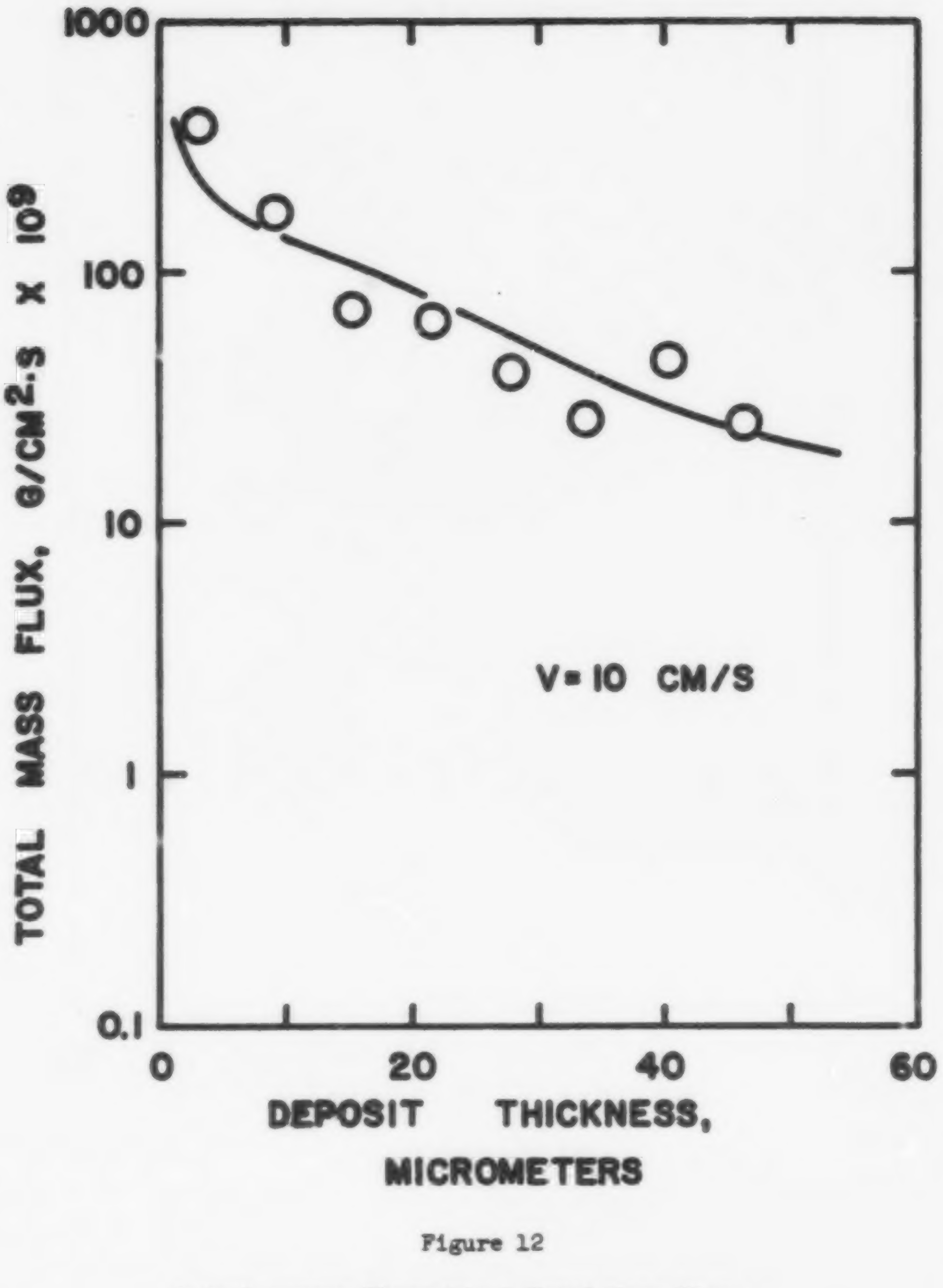

Figure 12

Actual versus Theoretical Total Mass Flux,

Velocity of 10 cm/s

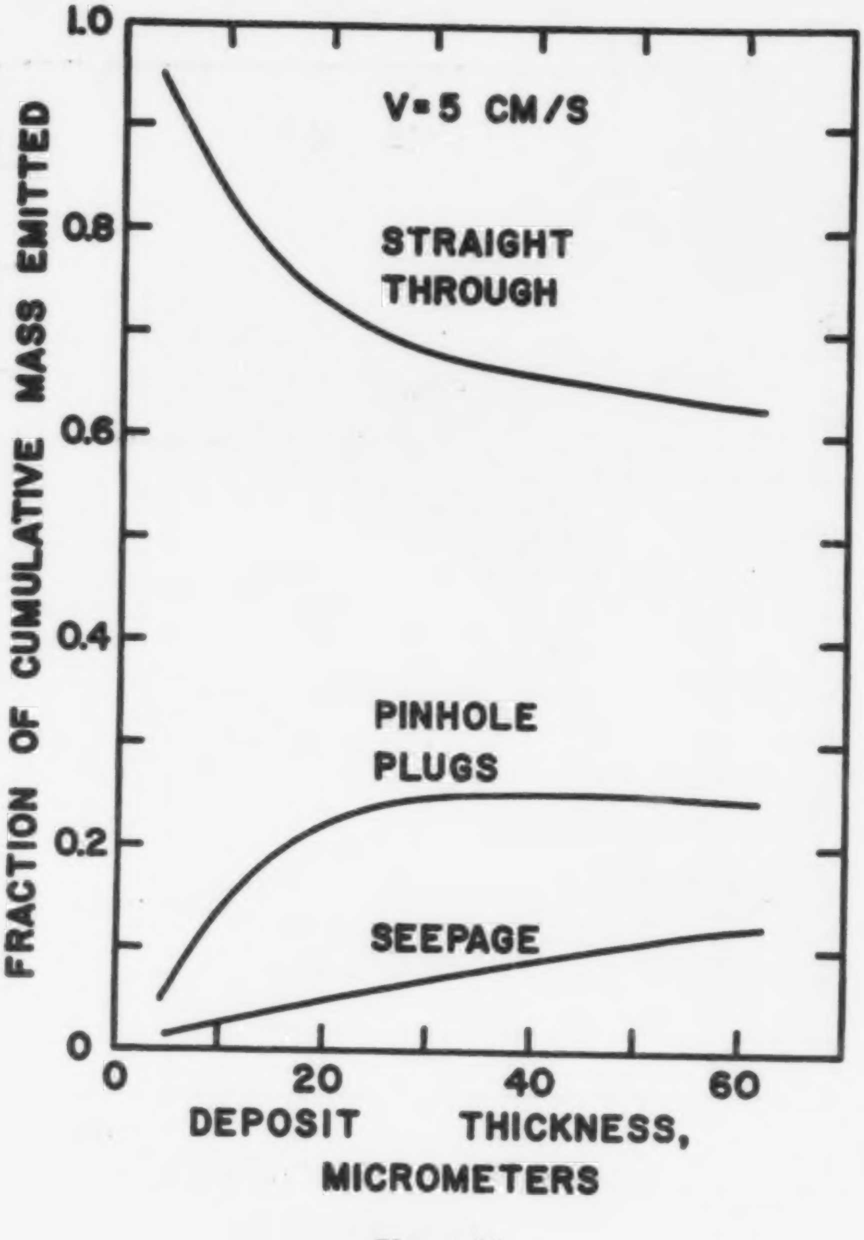

Figure 13

Fraction of Cumulative Mass Emitted Over Many Cleaning Cycles versus Deposit Thickness at Cleaning with Mechanism as Parameter, Velocity of 5 cm/s

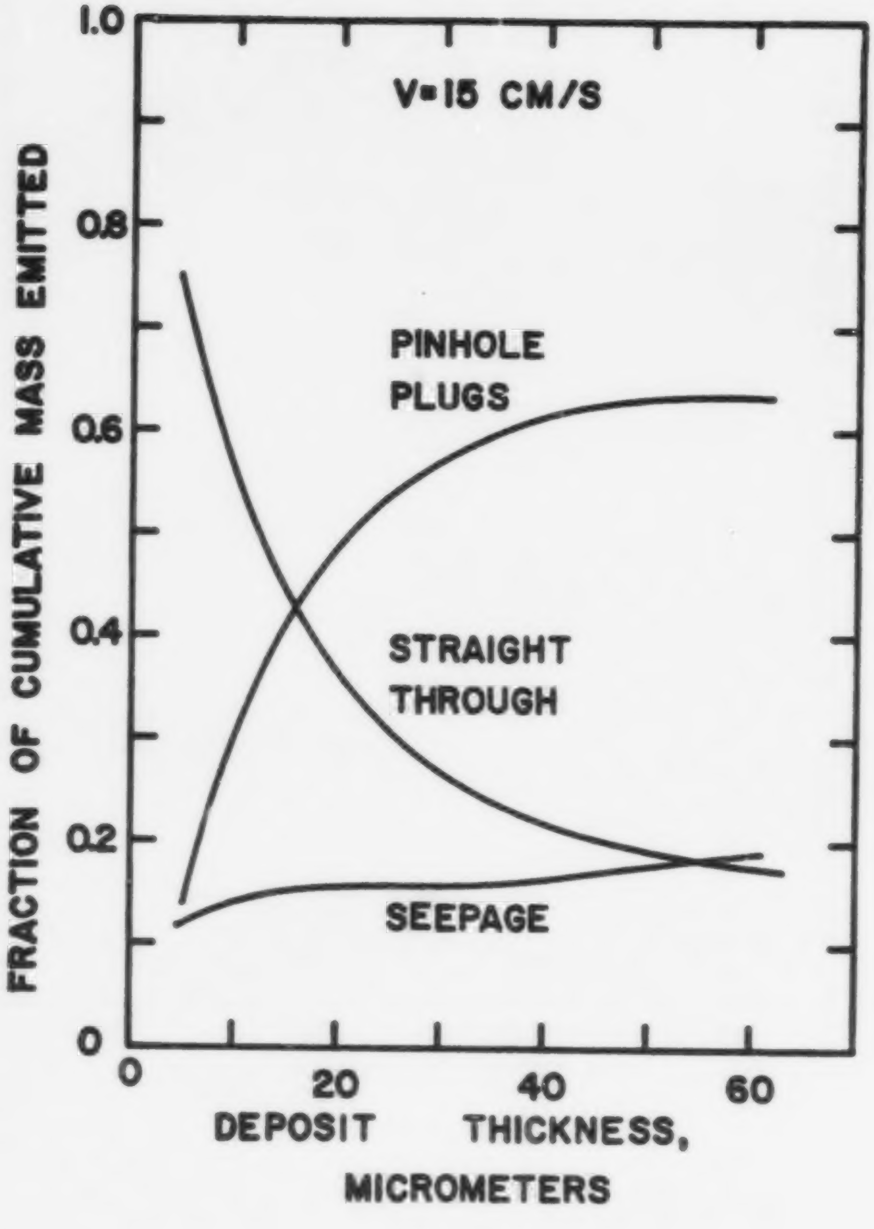

Figure 14

Fraction of Cumulative Mass Emitted Over Many Cleaning Cycles versus Deposit Thickness at Cleaning with Mechanism as Parameter, Velocity of 15 cm/s pinhole plug mechanism is largest, if cleaning does not occur until the deposit thickness reaches at least 15 micrometers as would normally be the case. The overall importance of the straight through mechanism decreases with increasing velocity, whereas that of pinhole plug generation increases.

Summary

The particle collection characteristics of a pulse-jet fabric filter collecting flyash have been studied. After cleaning, penetration and outlet mass flux decrease rapidly, but then level off. Penetration and outlet flux are nearly constant for all particle diameters, but decrease with decreasing face velocity through the filter bags. Pressure drop in excess of 100 cm of water prevented operation at velocities higher than 15 cm/s.

The rapid decrease in outlet mass flux after cleaning is coincident with a decline in the importance of the straight through dust emission mechanism. Thereafter, the outlet flux is controlled by the pinhole plug and seepage mechanisms. After a rapid rise, the pinhole plug mechanism allows outlet loading to decrease slowly as the dust deposit thickens. The outlet flux due to the seepage mechanism is relatively constant throughout the filtration cycle. The importance of the pinhole plug mechanism increases with increasing velocity through the fabric, whereas straight through penetration becomes less important.

Note:

Details of the data gathering procedure for penetration measurements, a description of the method develped for calculating confidence intervals about penetration values, details of the data analysis for the mechanism experiments, the theoretical justification for Equation 7, and a description of how the equations describing outlet mass fluxes were developed are all given in Reference 45.

Note on Size Selective Tag Deposition

The objective of the tagged dust experiments was to find out more about the mechanisms by which particles pass through a fabric filter. In designing these experiments, the results of the particle size versus efficiency experiments were available. Through these experiments, the particle size distributions upstream and downstream of the fabric filter were found to be virtually identical over all conditions of velocity and deposit thickness to be tested in the tagged dust experiments. This is shown by the forty independent size distribution measurements given in Tables A-4 through A-8 in Appendix A of Reference 45.

One should expect that size selective tag deposition occurs during the preparation of the tagged dusts, due to differences in chemical composition and surface area of the flyash with changing particle size. Chemical binding and surface adsorption are two means by which the tag molecules might be expected to join with the flyash particles. Because of the chemical similarity of the tag molecules and the uniformity of the flyash used, the size-deposition profile for all of the tagged dusts should be roughly the same.

However, due to the equivalence of the upstream and downstream size distributions, any size-tag bias present upstream would also be present downstream. In the same way that the analytical response to a sample of upstream dust indicates the mass of dust there, the analytical response to a sample of dust collected downstream will show the amount of dust there. Size selective tag deposition, therefore, was irrelevent to the design of the tagged dust experiments, and to the interpretation of the data generated.

If the size distributions of the upstream and downstream dust had not been similar, then size selective tag deposition would have made the quantification of downstream mass by tag analysis less direct. But because this size distribution similarity was known, the tagged dust experiments were planned and their results interpreted with confidence on this point.

Nomenclature--with units from the cgs system

c	aerosolized dust concentration, g/cm^3
f	fraction of dust emitted accountable to subscripted mechanism, dimensionless
K	constant, for dimensions see Table 5
t	time since last fabric cleaning, s
v	superficial filtration velocity, cm/s
X	dust deposit thickness, cm
z	fraction of total mass released over many filtration cycles accountable to one emission mechanism
ρ	flyash bulk density

Subscripts

e	aerosol entering the fabric filter
i	pertaining to any of the three emission mechanisms
o	aerosol coming from the fabric filter
pi	pinhole plug emission mechanism
se	seepage emission mechanism
st	straight through emission mechanism

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TOPICS FOR FURTHER CONSIDERATION

The study of fabric filters described here has enlightened areas which previously were speculation. As a result, many additional questions have arisen. A continuation of the work begun here will help find the answers to these questions, and is important if knowledge of the underlying factors governing fabric filtration is to progress.

One of the objectives of this work was to study the effect of high velocity operation on filter efficiency. Although pressure losses for the filter when operated at conventional velocities are low, velocities in excess of 15 cm/s caused high pressure drops, exceeding 100 cm of water column. There may be ways to mitigate this rapid pressure rise. The interplay of fabric construction with dust size distribution may play an important role in determining the speed with which pressure drop increases with increasing velocity. It might be that a less porous fabric or larger particle sizes would hinder particle deposition within the cloth, and help prevent possible "blinding" of the fabric with dust. A fibrous precoat might help by preventing the particles from entering the fabric structure to the extent that may presently occur.

In this work, it was found that dust passes through the fabric filter by straight through penetration, by seepage, and by pinhole plug loss. The sequence in which these mechanisms gain or lose relative importance with increasing deposit thickness was found to be relatively invariant with velocity. However, parameters not examined here may also be important. Relative humidity, particle shape, and fabric construction as well as gas, fabric, and particle physical properties may all play a role that is presently unknown.

Pulse-jet fabric filters currently make up the majority of the dollar sales for such dust collectors. Other cleaning methods are also used. The effect of velocity, deposit thickness, and particle size on efficiency for filters utilizing other means of cleaning should be studied, as well as the mechanisms by which these filters release particles to the atmosphere. Now that the techniques for making such studies have been developed, it would not require much additional developmental work to investigate other cleaning mechanisms.

Felted and woven fabrics are used in commercial fabric filters. The felt tested here is only one of a large number of similar felts. Fiber shape and diameter, staple material, needling intensity, felt porosity, and felt thickness may all influence filter performance. Woven fabrics differ among themselves in weave, staple fiber, finish, and other characteristics. Although the number of parameters involved in a study of fabric construction is large, it is necessary to understand the influence of each to acquire a complete understanding of all factors influencing fabric filter performance.

The importance of sub-micrometer particles is becoming ever more apparent as their adverse health effects become better known. In the present study, it was possible to determine efficiency for particles only as small as 0.27 micrometers diameter because of the limitations of the optical analytical method used. It is important to determine efficiency for particles smaller than this over ranges of velocity and deposit thickness similar to those used in the present study. Many commercial dusts are larger in diameter than 4.32 micrometers, the largest flyash particles observed regularly here. Efficiency as well as mechanisms of emission may change for much smaller and much larger particles, and these should be studied further.

The sites of pinhole plug generation may be locations at which the felt was needled at the time of its manufacture. It is important to establish whether this is the case. If so, then the manner and intensity of needling during the manufacturing process may profoundly influence the particle collection capabilities of the fabric filter, especially when the filter is operated at high velocity, a mode that emphasizes dust emission by the pinhole plug mechanism.

Modeling performed in this work assumed complete independence among the three dust emission models proposed. For example, dust which passes through the filter by the seepage mechanism was assumed to progress through the filter exclusively by this means. However, particles may pass through the filter by a combination of mechanisms. A particle might nearly pass through the filter by the straight through mechanism, and be caught near the clean fabric surface. Later, the same particle might continue its journey through to the clean side of the fabric by the seepage mechanism. Quantification of the interrelationships of the three emission mechanisms should provide an area for further research.

Finally, additional mechanisms, unidentified at present, may play a role in the release of dust to the atmosphere from a fabric filter. Should such mechanisms come to light, they can be studied through the approach outlined in this work. Such information would be vital to a further understanding of the way in which fabric filters function.

Section V

HIGH VELOCITY CAKE FILTRATION

GENERAL

Fabric filter technology has progressed only slightly compared to other fields of engineering endeavor. This, in part, is due to priorities set by society and in part because of the nature of the fabric filter industry. There are in excess of fifty manufacturers of cleanable fabric filters with an estimated annual 1970 sales of about fifty million dollars(1). The modest R & D effort of these manufacturers is directed primarily toward specific, application-oriented problems and in many cases remains proprietary. Basic experimental and theoretical research in this area has been limited in scope. There has, however, been considerable work relating to packed, granular beds and to non-cleanable, fibrous filters (e.g. filters used in air conditioning), and numerous attempts have been made to apply results to cake filtration. These attempts have met with only limited success. It is clear that research into the parameters controlling cake filtration itself is essential to produce improvements in cleanable fabric filters.

Cleanable fabric filters have normally operated at filter velocities ranging from 1-3 feet per minute (fpm) so that maximum pressure drop across the filter would be limited to 4 to 5 inches of water. In the last twenty years, the continuously cleaned reverse-jet filter and the pulse-jet cleaned filter were introduced, and these developments permitted filter velocities ranging from 10 to 20 fpm at essentially the same pressure drop. A maximum pressure drop of 4-5 inches of water was consistent with practices common in the past and reflected the low airflow resistance characteristics of the cyclone and the electrostatic precipitator. Recently, the Venturi scrubber and other versions of the high energy scrubber have made pressure drops as high as 70 inches of water common.

Since cleanable fabric filters have many exceptional properties, including simplicity, reliability, and collection efficiencies substantially higher than those of the electrostatic precipitator or high energy scrubber, it would be desirable to modify their operation to make them more compact and economical. Undoubtedly, a substantial increase in filtration velocity offers the greatest probability of a significant change in filter acceptability among users. To make high velocity fabric filtration a reality will require improvements in fabrics, possibly changes in cleaning techniques, and a basic knowledge of filter cake formation and performance.*

BACKGROUND

Fabric filter performance can be assessed by considering five interrelated factors:

(1) Pressure drop
(2) Collection efficiency
(3) Dust holding capacity
(4) Useful life
(5) Total cost.

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*Much of what is contained in this introduction is based on a paper given by Dr. M. W. First in 1971 at the Fabric Filter Symposium (2).

Cost and useful life, while essential to overall filter performance, are not of primary concern in this study. The primary objective of this work is to compare pressure drop, collection efficiency, and dust holding capacity at conventional filtration velocities (<20 fpm) and high velocities (>20 fpm) to provide the input necessary for an evaluation of high velocity filtration feasibility. The study will also provide a basic understanding of cake filtration at conventional velocities. The following background material is pertinent to these ends.

Pressure Drop - An increase in filtration velocity should result in a rise in pressure drop across the filter cake as given by D'Arcy's law. The pressure drop across the clean cloth is not considered here since it is negligible compared to that of a well-formed cake or even a particle saturated cloth. That is not to say that the choice of fabric is unimportant. The fabric has a profound effect on the pressure drop across the filter cake and across the saturated fabric as demonstrated by the work of Spaite & Walsh(3) and Durham & Harrington(4). D'Arcy's law is given by the following equation:

$$\Delta P = \frac{\mu V L}{K_d} \qquad (14)$$

where ΔP = Pressure drop
μ = gas viscosity, poises
L = thickness of the cake, cm
V = approach velocity, cm/sec
K_d = D'Arcy permeability, cm^2

It is instructive to examine the assumptions implicit in equation (14) and how they are affected by higher velocities. The important assumptions are: (1) The pressure drop across the bed is small compared to the total pressure (i.e. incompressible gas), (2) The resistance to flow is controlled by viscous and not inertial forces (i.e. laminar flow), and (3) there is no gas "slip." The first assumption is easily satisfied for conventional fabric filtration where the pressure drop is about 5 inches of water (0.01 atmospheres), a small value compared to atmospheric pressure. However, if the filter velocity is increased such that the pressure drop is 100 inches of H_2O (0.25 atmospheres), the gas will expand by roughly 25% after flowing through the filter, and hence the linear velocity will increase by that same amount causing an additional rise in pressure drop. D'Arcy's equation when corrected for this effect takes the following form:

$$\Delta P = \frac{P_1}{P_{ave}} \frac{\mu V L}{K_d} \tag{15}$$

where P_1 = inlet pressure, dynes/cm^2

P_{ave} = average pressure, dynes/cm^2

The assumption of laminar flow can be tested by computing the Reynold's number from the following equation:

$$Re = \frac{VD_p}{\nu(1-\varepsilon)} \qquad (16)$$

where Re = Reynold's number, dimensionless

V = superficial velocity, cm/sec

ε = void fraction of deposit (porosity)

D_p = Sauter mean diameter*

ν = gas kinematic viscosity, cm^2/sec.

If a value less than about one** is obtained, the flow resistance is primarily viscous, and D'Arcy's law is applicable. (This should be experimentally verified for filter cakes). To determine the superficial velocity at which inertial forces begin to play an important role, the Reynold's number can be set equal to one in equation (16), and solved for V. Assuming a Sauter mean diameter of 10μM, a corresponding fractional void volume of 0.7*** and normal temperature and pressure, the calculated superficial velocity is approximately 100 fpm. If the particle size is less than 10μM diameter or if the temperature* is increased, the velocity at which transition flow begins will increase above 100 fpm.

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*Diameter of a sphere whose volume/surface area ratio is the same as that of the dust cake.

**This is probably conservative. Silverman(5) and Leva(6) give Re < 10 for laminar flow, and Happel & Brenner(7) give Re < 5.

***Approximated from reference (8).

If the Reynold's number exceeds one, another term must be added to D'Arcy's equation to correct for inertial effects as shown below:

$$\Delta P = \frac{\mu LV}{K_d} + \frac{\rho LV^2}{K'} \qquad (17)$$

where ρ = density of the gas, gm/cm^3

K' = constant, cm (depends on the same parameters as D'Arcy permeability).

This equation applies to both the transition and turbulent flow regions, but is is unlikely that fully turbulent flow will ever be approached in cake filtration. Fully turbulent flow would correspond to about 10,000 fpm** based on 10μM particles and N.T.P.

The validity of assuming no gas "slip" depends on the size of the particles relative to the mean free path of the gas. When they are of the same order of magnitude a correction must be applied to equation (1) as follows:

$$\Delta P = \left(\frac{P_1}{P_{ave} \mp B}\right) \frac{\mu VL}{K_d} \qquad (18)$$

where B = a constant dependent on gas and cake properties

---

*If the temperature is increased from 20°C to 100°C, the calculated velocity becomes 200 fpm.

**Re > 100$^{(5),(6)}$

It should be noted that inertial effects and gas "slip" can never occur simultaneously.

Assuming equation (14) is valid, the only parameters affected by an increase in velocity other than pressure drop are the cake thickness and D'Arcy permeability. Equation (14) for D'Arcy's law can be modified to:

$$\Delta P = \frac{\mu V W}{K} \tag{19}$$

where W = areal density*, gms/cm^2

K = modified D'Arcy permeability, gm/cm**

The areal density will, of course, be influenced by the velocity since it will increase more rapidly with time due to the higher particle flux, but this effect could, presumably, be controlled by variation of the cleaning cycle. Therefore, it is more useful to observe the cake at a given instant in time such that the above effect will be eliminated.

A relationship delineating permeability derived by Kozeny(9) and modified is:

$$K = \frac{\rho_t \, \varepsilon^3}{kS^2(1-\varepsilon)} \tag{20}$$

where k = Kozeny constant***

S = specific surface (surface area of particles per volume of particles), cm^{-1}.

ρ_t = true density of the dust in the cake, gm/cm^3

---

*Weight of dust deposited per unit area of cloth

**Henceforth referred to as "permeability"

***Theoretically equal to 5 for spheres(9) or experimentally equal to 4.2 from a large amount of data(10).

This equation yields excellent results for powders which are compressed to a specified porosity provided the particles are isometric. It is based on a model involving follow through parallel passageways. If the pores are arranged in parallel, the larger pores would be expected to carry a significantly greater portion of the flow than the smaller pores in accordance with the Hagen-Poisueille relationship. However, since for filter cakes it is the flow itself which creates the porous media, it would be expected that many particles would be brought to the mouth of a large pore as a result of the high flow rate (and in particular the smaller particles which tend to flow streamlines). Therefore, it is likely that large pores would connect with small pores (i.e. pores in series) thereby decreasing the permeability of the porous media relative to a model in which it is assumed that all pores are in parallel. Hence, equation (20) may not be adequate for filter cakes except qualitatively.

It can be observed from equation (20), that porosity appears as a complicated function of great sensitivity. Unfortunately, as noted in the Handbook of Fabric Filter Technology, "The actual porosities of dust deposits on operating fabric filters have been measured." They estimate the probable range of porosities encountered will vary from 0.4 to 0.9(11). This corresponds to nearly a 70-fold change in permeability. The porosity of the cake depends on many interrelated variables including:

(1) Particle size distribution

(2) Particle shape and surface roughness

(3) Adhesion and frictional forces

(4) Humidity

(5) Electrostatic effects

(6) Pressure drop across the cake

(7) Drag forces on the particles in the cake

(8) Deposition velocity for incoming particles.

Since the major interest is a comparison between low and high filter velocities, many of the above parameters can be eliminated from consideration (at least to a certain extent). For example, if the same aerosol cloud is employed, the particle characteristics will not play a direct role in the comparison. Similarly, effects due to humidity and electrostatics can be minimized by proper experimental techniques. The parameters that remain, then, are those which are dependent on the velocity itself. Higher velocities can cause changes in cake porosity primarily in four ways:

(1) By increased small particle deposition within the cake

(2) By tighter particle packing at the surface of the cake

(3) By compaction of the cake

(4) By cake punctures

Increased small particle deposition within the cake should be related to small particle collection efficiency which will be discussed later in this section.

Tighter particle packing is due to the larger momentum exchange of particles landing on the surface of the cake. This is probably a significant effect which should be easily extractable from experimental data.

Compaction of the cake is a complicated phenomena which is probably due to the increased drag exerted by the gas on the cake. Although the dust on the surface of this cake is under negligible mechanical stress due to this drag, the dust below the surface must support the drag experienced by the dust layer above it. At the bottom of the cake (the dust-fabric interface) the compressive stress is the greatest and equals the total pressure drop across the cake, plus the stress due to the landing of new particles. This implies that the nature of the dust-fabric interface (and hence the fabric) are of great importance since the projecting fibers of the fabric aid in supporting the cake at the point at which the compressive stress is greatest. This suggests the use of a heavily napped cloth with stiff, relatively long projecting fibers, although these fabrics are more difficult to clean.

Limited studies have been done to determine the effects of increased velocity on permeability.* Stephan, et al(12) observed a 60% change in this factor (including a 50% reduction in cake thickness) when velocity was increased from 6 to 10 fpm with clean air, and the change was noted to take place in 5 discrete collapses. The cake was initially deposited at 3.4 fpm. Borgwardt, et al(13) reported a similar change with the same dust (fly ash). These studies provide some insight into cake collapse. However, if the cake is formed at high velocity, and the velocity remains constant, the compaction of the cake may proceed more gradually due to tighter packing initially.

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*Stephan, et al(12) used the factor K/μ which they called dust permeability. Since the gas viscosity is constant for the above discussion, dust permeability is directly proportional to permeability.

Compaction can take place by three mechanisms(14): (1) sliding, (2) elastic and plastic deformation, and (3) fragmentation. The latter is unlikely except for extremely fragile particles due to the relatively small compressive stresses. The most likely mechanism is sliding which is opposed by frictional and cohesion forces between particles in the cake. Compaction effects in powder beds are often described by exponential expressions such as the following(14):

$$\varepsilon_f = \varepsilon_i \text{ EXP } (-aF) \qquad (21)$$

where ε_f = final void fraction (porosity)

ε_i = initial void fraction (porosity)

a = constant, cm^2/dynes

F = compressive stress, $\frac{dynes}{cm^2}$

Its applicability to cake filtration has never been demonstrated. Cake compaction may also be related to flexing of the support medium due to the pressure drop.

Cake puncture is another problem which is expected at higher velocities. This is conceived as a collapse of the cake very close to a high velocity spot in the cloth(12), and the subsequent ejection of a tiny piece of cake through the cloth. Punctures are most prevalent with coarse cloths and fine dusts. Stephan et al(12) report that punctures could not be induced even at very high pressure drop when a cake support of low porosity (glass filter paper, membrane filters, etc.) was employed. However, with cotton sateen cloth as the cake support, punctures occurred at 10 fpm. The possibility of cake compaction and cake punctures suggests the importance of a low porosity cloth and a rigid support backing for the study of cake formation at high velocities.

Collection Efficiency - Insight into the effect of high filtration velocities on dust retention can be obtained from a study of classical filter theory(15) developed primarily for high porosity fibrous filters. The mechanisms of particle capture in fibrous filtration and cake filtration are the same. They are:

(1) Inertial deposition
(2) Deposition due to Brownian diffusion
(3) Gravitational settling
(4) Deposition due to electrical effects
(5) Direct interception
(6) Sieving.

The latter is an important characteristic of cake filters only.

Two assumptions are generally made in classical filter theories: They are: (1) the collecting obstacles are sufficiently far apart so that the fluid flow in the vicinity of an obstacle approaches the flow around an isolated obstacle, and (2) the particles adhere on contact and are then fixed (i.e. the particles do not bounce, migrate once attached, or become resuspended). The first assumption, though reasonable for many fibrous filters, is not valid for cake filtration. The second assumption may not be valid for cake filtration depending on operating conditions. In a gross way the phenomena of cake collapse, cake puncture, and dust seepage are violations of this second assumption. Although not directly applicable, classical filter theory will provide insight for modeling the mechanisms involved in cake filtration.

Recently, Paretsky, et al(16) proposed a semi-theoretical prediction of aerosol capture by Brownian diffusion, inertial impaction, gravity settling and direct interception for granular beds based on Happels "free cell" model. This model has been successfully employed to predict the Kozeny constant of equation (7) starting from the Navier-Stokes equation(9). Despite this success, the model has some limitations for cake filtration. The most obvious one is the absence of a sieving mechanism in the model. (Their prediction for direct interception will not yield valid results for sieving). This omission is defensible in the context of their experimental work (and that of Thomas and Yoder(17)) on packed beds of sand which they call on for

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verification of their model since sieving of the aerosol particles employed (<1.1μm diameter) is unlikely. For this model the experimental data agree reasonably well for diffusional collection in the range where diffusion is the dominant mechanism of aerosol capture. However, the results for inertial impaction are contradictory. The authors note that the "discrepancy between calculated and experimental results is not now understood." The experimental and theoretical results for gravitational deposition are within an order of magnitude. No verification for direct interception could be obtained due to its relative unimportance. Calculation of total collection efficiency from theoretical individual collection mechanisms remains an unsolved problem.

Attempts to refine and extend theoretical efforts should also take other factors into consideration. Higher velocities might enhance forces tending to cause re-entrainment and particle migration. On the other hand, the additional compressive stress on the particles and increased contact area, due to the higher velocity, might enhance cohesion and frictional forces between particles in the cake. In the studies of Wright et al(18) in which deposited particles were subjected to successively higher air velocities, substantial re-entrainment occurred only at velocities of about 2000 fpm.

Since a well-formed cake captures a large percentage of particles by sieving, depending on the dispersity of the aerosol cloud and its packing characteristics, higher velocity would not be expected to hinder this mode of collection. In fact, if the higher velocity caused cake compaction, as is likely, the minimum particle size retained by sieving may be reduced. However, even when the smallest particles in the aerosol cloud are capable of passing through the pores of the filter cake, they are susceptible to retention by classical filtration mechanisms. Therefore, the primary concern in evaluating the effect of high filtration velocities on dust retention should focus on (1) the overall efficiency of saturated fabrics during the brief periods following cleaning when the new filter cake is forming, and (2) the efficiency of the cake for particles too small to be removed by sieving. For the latter case, when the particles are sufficiently small, they will be removed primarily by diffusional forces, and their penetration can be predicted from the model of Paretsky, et al$^{(16)}$as follows:

$$n_\rho = 5.04 \left[\frac{1 - (1-\varepsilon)^{5/3}}{2-3(1-\varepsilon)^{1/3} + 3(1-\varepsilon)^{5/3} - 2(1-\varepsilon)^{2}}\right] \left[\frac{\mathcal{D}}{VD_p}\right]^{2/3} \qquad (22)$$

where n_ρ = single collector efficiency for diffusion

ε = void fraction of deposit (porosity)

$\mathcal{D}$ = diffusion coefficient of particle, cm^2/sec

V = superficial velocity, cm/sec

D_p = Average collector diameter (particle size in cake), cm

The single collector efficiency is easily relatable to total efficiency provided it is the only collection mechanism considered. Equation (22) implies that the efficiency attributable to diffusion will decrease as velocity is increased. However, since the complicated function of ϵ in the above expression increases with decreasing porosity, the decrease in collection efficiency noted above will be offset by an increased efficiency due to cake compaction. It would be surprising, though, if the decreased residence time due to high velocity and reduced porosity, could be fully compensated for by changes in the compacted cake (i.e. decrease in deposition distance and a higher tortuosity*). At higher velocities, increased efficiencies will also be observed because smaller particles will be more prone to inertial collection. This trade-off of more inertial retention and less diffusional collection is likely to increase the total efficiency based on mass since the particles removed by inertia are heavier than those removed by diffusion.

Summary of Experimental Work

The work done thus far primarily concerns the effect of velocity on cake filtration. Using a representative dust and filter cloth, the relationship between cake formation velocity and cake properties (e.g. porosity, permeability, pore size and size distribution, ability to resist cake collapse) is being determined. The studies include the determination of pressure drop-time relationship during cake build-up, permeametric studies of the cake at various times during its formation, and microscopic studies of the overall cake geometry and cake pore structure. Hopefully, this work will aid in evaluating the feasibility of increasing velocity in industrial, cleanable bag filters. It will also allow the interrelationship between the various parameters involved in cake filtration to be better understood.

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*Defined as the ratio of the total distance a particle travels when penetrating the filter/filter thickness.

Much of the work thus far has involved the development of experimental techniques and equipment. A method has been developed to accurately determine the porosity of cakes supported on woven cloths. Techniques have been developed to view the pore structure of filter cakes using the scanning electron microscope. Equipment has been built and pilot runs made to form filter cakes at various operating conditions such that important parameters can be recorded as a function of time.

EXPERIMENTAL WORK

Equipment

Many modifications in the experimental apparatus have been made, and only the final design will be discussed fully in this section.

The use of a bag filter would present many problems: difficulty in analyzing the cake, dust elutriation in the bag, impracticality of studying various types of fabrics and fabric supports, and others. Therefore, bench-scale equipment similar to that of Williams et al$^{(19)}$ and Stephan, et al$^{(12)}$ was utilized. With this configuration, a flat piece of cloth is supported horizontally. Williams, et al$^{(19)}$ used circular swaths of cloth 6 inches in diameter whereas Stephan, et al used 1.5 inch diameter cloths. The exact size is not critical provided the ratio of particle diameter to cloth diameter is small, thereby making the edge effects unimportant.

Figure 15 is a schematic diagram of the experimental apparatus. Compressed air after being cleaned is fed to the Wright dust feeder in which a packed powder cake is continually scraped, aerosolized, and dispersed. A minimum pressure of 3 psia is required to adequately disperse the dust, and this pressure is maintained and held constant through use of a mercury manostat which allows the excess flow to be bubbled through mercury. The aerosol stream passes through a small chamber containing an Americium 241

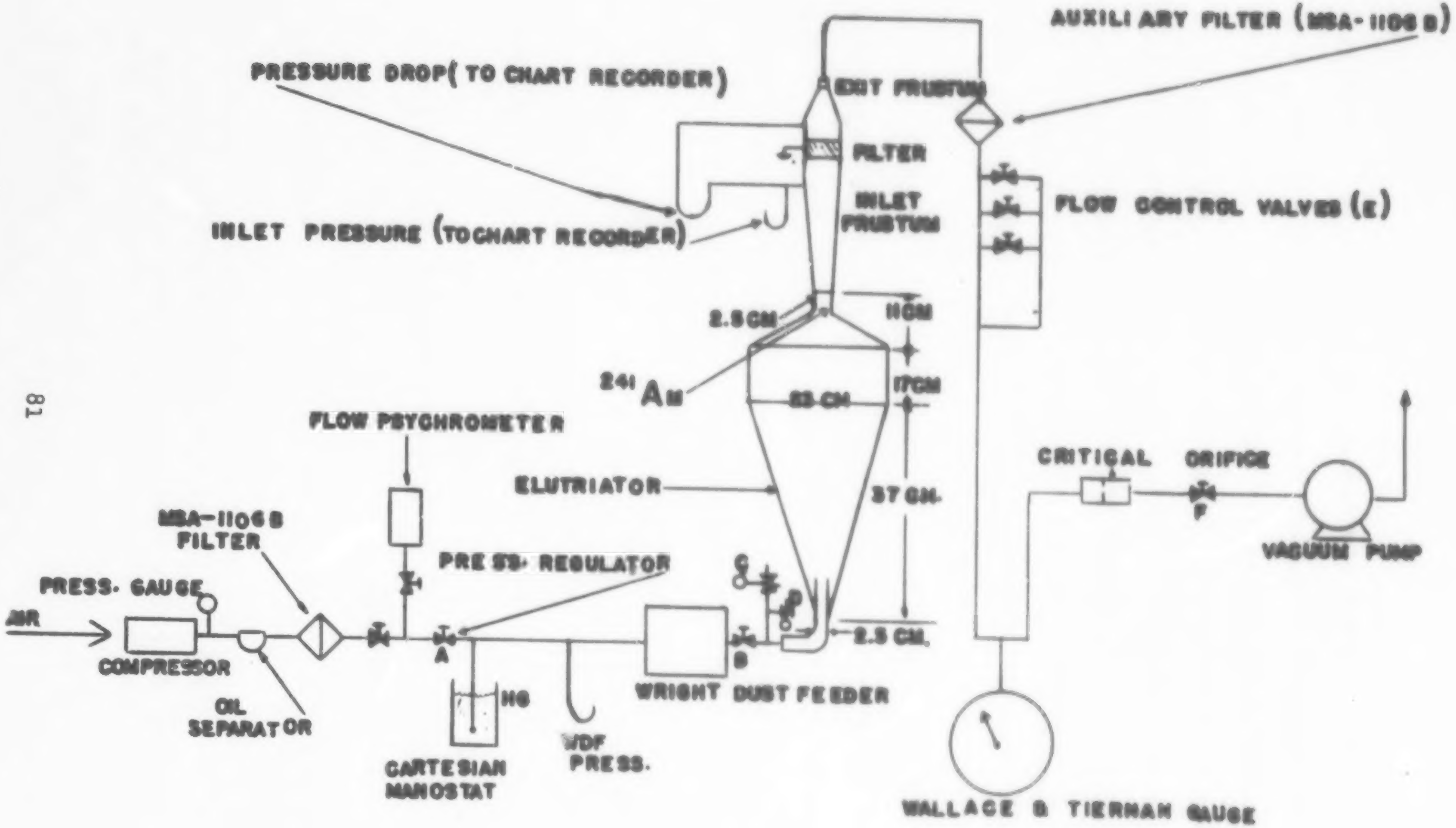
AUXILIARY FILTER (MSA-1106 B)
PRESSURE DROP (TO CHART RECORDER)
EXIT FRUSTUM
FILTER
INLET FRUSTUM
INLET PRESSURE (TO CHART RECORDER)
FLOW CONTROL VALVES (E)
2.5 CM
11 CM
^{241}Am
17 CM
23 CM
FLOW PSYCHROMETER
ELUTRIATOR
37 CM.
CRITICAL ORIFICE
F
VACUUM PUMP
MSA-1106 B FILTER
PRESS. GAUGE
AIR
COMPRESSOR
OIL SEPARATOR
PRESS. REGULATOR
A
C
D
B
2.5 CM.
WRIGHT DUST FEEDER
HG
CARTESIAN MANOSTAT
WDF PRESS.
WALLACE & TIERNAN GAUGE

Figure 15

Experimental Apparatus

source (0.1 mCi) in which the aerosol is neutralized.* This source prevents the build-up of a high residual charge on the test particles, a common result during dry powder dispersal. The stream then enters an elutriator in which all particles larger than 11 micrometers are removed.** The Reynolds number in the elutriator varies from 75 to 660, and hence the flow should be laminar. A reservoir at the bottom of the elutriator serves as a receptacle for the dust which is removed. The flow then passes upflow through the filter, a bank of valves which are used to control the flow rate (the increased resistance of the cake as it grows is compensated for by decreasing the downstream resistance through the control valve, thereby holding the mass flow rate constant) and a critical orifice.*** A Wallace and Tiernan gauge, which can measure pressure very accurately, is placed directly upstream of the critical orifice. A "Leiman" brand vacuum pump is used downstream of the critical orifice. The pressure drop across the filter and the pressure in the elutriator are recorded automatically on a chart recorder. The filter holder is grounded. The use of a critical orifice has the following advantages:

(1) Eliminates effect of pressure oscillations inherent in the vacuum pump.

---

*Charge equilibrium of the aerosol was calculated based on work of Cooper and Reist(20).

**The removal of all particles greater than 11μm in the elutriator was calculated using Stokes law and assuming laminar flow.

***Since a critical orifice yielding the required mass flow rate was not available, two critical orifices in parallel were used.

(2) Stabilizes flow rate by acting as a high flow resistance in series with the filter such that a small change in pressure drop across the filter cake does not have a significant effect on the mass flow rate.

(3) Can be used to control and monitor mass flow rate since the pressure upstream of the critical orifice is directly proportional to mass flow rate regardless of the downstream pressure. The critical orifice and vacuum pump combination was calibrated and the flow rate is given by the following equation provided the upstream pressure is greater than 275 mm Hg.

$$Q = 21.8P \tag{23}$$

where Q = flow rate, lpm at ambient conditions

P = upstream pressure, mm Hg.

During a run, as the cake builds up, the flow control valves are adjusted such that the pressure upstream of the critical orifice, as read on the Wallace and Tiernan gauge, remains constant, and hence the mass flow rate is constant. Simultaneously, the pressure drop is recorded on a chart recorder. The pressure upstream of the filter remains constant near atmospheric pressure and is also recorded automatically.

The approach velocity is adjusted by varying filter area. The apparatus is designed so that only the filter holder and connecting pieces require modification to change the filter area. All operating conditions except filter face velocity thereby remain unchanged. The minimum velocity which can be obtained is 1 fpm corresponding to the maximum available diameter of the elutriator.[*] The maximum velocity, limited by the onset of turbulence in the entrance section of the filter, is about 500 fpm.

The filter holder, an important part of the apparatus, is shown in Figure 16. It consists of a wire screen and filter cloth mounted between two concentric brass rings. Since the cloth is taut and supported by a nonflexible screen, stretching of the cloth is retarded. The entrance section to the filter consists of a cylindrical ring of the same diameter as the filter mated to a frustum with a maximum half-angle of 7° to eliminate boundary layer separation. The outlet section, though similarly constructed, requires only a 30° half-angle. The filter and connecting pieces are butted to each other, and secured with teflon and vinyl tape. This arrangement prevents leaks and allows the filter to be removed easily without upsetting the cake. The filter holder is sufficiently small so that the cake can be weighed, viewed microscopically, or infiltrated with plastic without being removed from the holder.

A flow psychrometer requiring low flow rates (~ 20 lpm) is connected to the air supply preceding the dust feeder. The relative humidity is held constant and controlled

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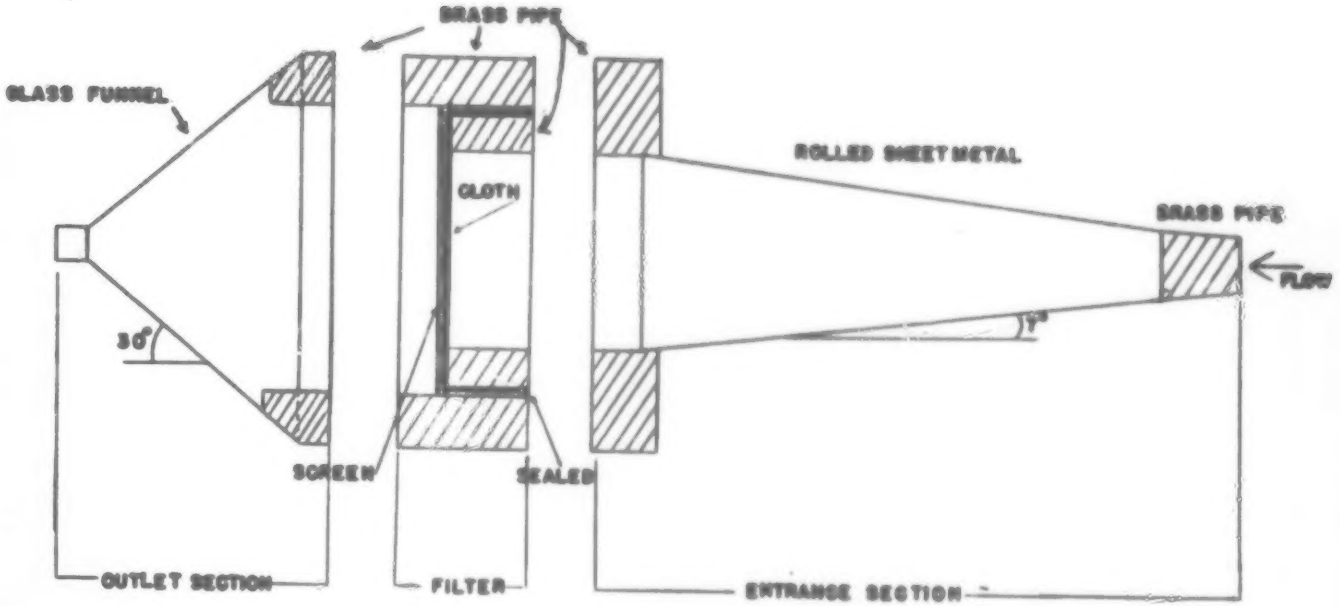
[*] If the filter has a larger diameter than the elutriator, the aerosol size distribution will be dependent on the filter size.

Figure 16


BRASS PIPE
GLASS FUNNEL
ROLLED SHEET METAL
CLOTH
BRASS PIPE
FLOW
30°
7°
SCREEN
SEALED
OUTLET SECTION
FILTER
ENTRANCE SECTION

Prototype Filter & Connecting Pieces
(Cross-Sectional View)

by adjustment of an air compressor.*

This set-up has the following desirable features during cake formation:

(1) Constant mass flow rate of gas with a minimum of adjustment.

(2) Constant pressure and velocity in the elutriator and at inlet of filter (and hence an invariant aerosol).

(3) Atmospheric pressure at the inlet of the filter.

Step-by-step instructions for cake build-up, velocity scans (ΔP versus flow rate), and permeametry at reduced pressure (ΔP at various flow rates and pressures) are given in Appendix H . These instructions have been followed successfully except for one particularly bothersome problem.

After the cake was allowed to build-up to some desired thickness, a velocity scan was made by halting the dust flow and incrementally decreasing the flow rate while noting the pressure drop until flow ceased. The original flow rate was then re-established and the pressure drop noted. It was always observed to be lower than before the velocity scan. Repeating the procedure produced another decline in pressure drop. The effect was more pronounced, the thicker the cake. It is believed that this decline in pressure drop is due to movement of the screen support caused by a changing stress on the screen equal to the pressure drop. The resulting fissures in the cake were observed with a microscope. This effect could not be observed while the cake is being formed because cracks caused by the screen movement heal themselves with fresh dust.

---

*The constant relative humidity system is very simple in principle. Ambient laboratory air almost always has a relative humidity in excess of 20%. If this air is compressed to 60 p.s.i.g., it becomes saturated, excess water being eliminated in the compressor. A large holding tank allows the saturated, compressed air to equilibrate at ambient temperature. Upon expansion to atmospheric pressure, which is assumed to be isothermal, the relative humidity will always be 20%. Compressing the air at lower pressure would increase the relative humidity.

To restrict movement of the wire screen, the following were tried: (1) a thumb screw to support the center of the screen, (2) a stiff spider to support the back of the screen, and (3) a ring concentric to the filter holder equi-spaced between the center of the screen and its circumference. The latter proved most effective, but still did not completely eliminate the problem.

At 50 fpm face velocity, punctures in the cake were numerous and could be seen with the naked eye. Using a microscope, the diameter was determined to be 200-300 μm. Punctures were eliminated by using two layers of cloth supported by a stiff wire screen. Punctures were not detected at lower face velocities (6 fpm).

Particle Analysis

AC fine Arizona air cleaner test dust was selected for all experimental runs to date. Analyses of this dust yielded the data shown in Table 6.

Table 6

Size Analysis for Arizona Road Dust

Method	Source	CMD	MMD	σ_g	Sauter Mean Diameter* (est.)
Sieving	Manufacturer		7.6µm	4.6	2.4µm
Elutriation	Peterson & Whitby(21)		5.4µm	4.1	1.9µm
Light Microscopy (450X)	S.N. Rudnick	1.5µm		2.6	

The smaller σ_g obtained by light microscopy may be due to the inability of light microscopy to detect particles smaller than about 0.3µm or to find the relatively few large particles that make a major contribution to mass. It may also be due to a nonrepresentative sample. (A water suspension of powder was dried on a microscope slide).

The density of Arizona fine road dust was determined by the method of air pycnometry**. The density is 2.8 gms/cm^3.

Cake Porosity Measurements

For modeling purposes, the porosity of filter cakes is a useful parameter. The actual porosities of dust deposits on operating fabric filters have never been measured(11). The porosity of a filter cake can be calculated from the following equation:

$$\varepsilon = 1 - \frac{W}{L\rho_t} \tag{24}$$

---

*Diameter obtained by dividing volume of particles by surface area. This is the diameter used in the Kozeny-Carman equation.

**A weighed sample of powder is placed in a calibrated glass container, and the entrapped air is compressed by displacement with mercury. From the change in pressure, the volume of powder (and hence the density) can be determined.

where ε = porosity

W = areal density, gm/cm^2

L = cake thickness, cm

ρ_t = true density of dust in the cake, gm/cm^3

The areal density and density of the dust are known or easily measured.

Two techniques were developed to determine the thickness of filter cakes. For those cakes deposited on filter paper, a fracture was made and a portion of the cake removed. Using a microscope with overstage illumination, the surface of the paper was brought into focus and a reading was taken on the micrometer dial of the fine focus adjustment. A similar procedure was followed for the surface of cake. A standard microscope with almost any light source providing side illumination proved to be adequate. With a 10X objective, errors on reproducibility of as much as 10% are present. At higher magnification (a 22X objective was the most powerful objective which allowed sufficient lighting) reproducibility was within about 3%.

The above described method was not adaptable to cakes supported on cloth filters because the surface of the cloth was not flat. Therefore, an alternate method was developed. The procedure involved the removal of a plug from the cake with a blunted hypodermic needle connected to a vacuum source without disturbance of the surrounding portions of the cake. Using a steroscopic microscope, the height of the plug is measured with the cake tilted at various angles. This yields the volume of the cake above the uppermost surface of the cloth. The volume of cake below the top surface of the cloth is measured independently by embedding the clean cloth in plastic, and then making microscopic measurements.

Microscopic Analysis of Cake Cross-Sections

Experimental equipment was assembled so that a dust cake could be infiltrated with a liquid without disruption of the cake. The dust cake is evacuated and then carefully flooded with an uncured plastic. The dust cake has a tendency to float, but this can be prevented by proper geometry. Upon repressurization the pores of the cake fill with plastic. The plastic is then allowed to harden at as slow a rate as practical. Various plastic materials* were tried although only "Clear Cast" proved adequate of those plastics evaluated.

The infiltrated dust cake is then smoothed with increasingly finer grades of wet sand paper.** Final polishing is done successively with a aqueous suspension of 0.3µm and 0.06µm alumina particles sprayed on a rotating napped cloth. Care must be taken particularly in the latter stages of this procedure to remove all surface irregulatities caused by the previous polishing steps.

The viewing of the polished stabilized dust cakes must be done with a microscope having vertical illumination(22). A standard microscope using a Bausch & Lomb vertical illuminator and an inverted microscope* designed for metallurgical work both gave good results although glare was somewhat of a problem at higher magnifications. Since the dust cake contained many particles below 1.0µm, these particles were not easily resolved. This difficulty was eliminated by using a scanning electron microscope. An extremely clear photomicrograph of a cross-section of cake at magnifications up to 5000X** was obtained. The structure of the support medium was also well shown.

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*Glycol methacrylate, "Ciba" epoxy resin, "Clear Cast" polyester resin.

**80, 120, 240, 320, and 600 grit.

Photomicrographs of the stabilized and polished cake indicated only minor changes in the cake from its assumed structure, The obvious artifacts include the following: (1) Debris on the polished surface, which judging by their size, are primarily the alumina polishing particles. (2) Gaps at the edges of the larger particles probably caused by shrinkage of the plastic material on curing. (3) The particles raised relative to the plastic material, probably due to the difference in hardness of the two substances. (4) Darkened spots on the photomicrographs probably the result of particles being "plucked" during the sanding and polishing steps.

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*American Optical Metallograph

**The scanning electron microscope has the capability of 100,000X magnification with a resolution of 0.01µm.

An early objective of this work was to develop techniques such that a cross-section of cake parallel to the director of flow could be viewed microscopically. Porosity could then be determined as a function of distance from the surface of the cake, thereby providing means of quantitatively evaluating compaction of the dust cake. Another reason for studying cross-sections of dust cakes microscopically was to evaluate the geometric structure of the cake and particularly the pore size distribution.

The techniques are now available for producing photomicrographs of cake cross-sections and a limited number have been made. However, translating those photomicrographs into useful numerical quantities still remains problematic. The cake cross-section appears as a collection of flattened particles surrounded by open space. This open space is, of course, the pores although it is more analogous to a large open area with many, small discrete obstacles dispersed throughout. Two questions arise:
(1) How does one define pore size (and size distribution)?
(2) Is this definition of any real use? As pointed out by Scheidegger(23), in the literature people write about the "size" of pores, "pore size distribution," etc., without defining accurately what this means. He proposes a definition as follows: "The pore diameter at any one point within the pore space is the diameter of the largest sphere which contains this point and remains wholly within the pore space." This could also

specify pore size distribution. This definition, with some modification, would provide a rather straightforward, though time consuming, method for determining the pore size distribution from photomicrographs. If a set of random coordinates were selected for each photomicrograph, the diameter of the largest circle containing the given point and inscribed in the open space could be called the pore diameter for that point*. Other variations on this scheme are also possible.

The second question involving the application of pore size distrubution is more difficult to solve. A common belief in the field of flow through porous media is expressed by Carman(24) as follows: "Large capillaries give disproportionately large rates of flow and so more than counterbalance the loss for small capillaries. In fact, if non-uniformity is very marked, flow through large capillaries alone is more than the total rate for uniform capillaries, and the volume and surface of the smaller capillaries play so little part that they can be ignored altogether." This is really just a result of Poiseuille's law which states that the flow rate through a capillary is proportional to the capillary diameter to the fourth power. The above belief is, of course, true if a porous medium is a collection of capillary tubes of various sizes arranged in parallel. However, in reality, the pores are arranged in parallel and in series. If all pores were in series, it would be the smallest pores, not the largest, which determine the resistance to flow through the porous media. Hence, information on pore size distribution is not likely to be useful unless it is known how the pores are interconnected. It does not seem possible to get this information from a 2-dimensional photomicrograph.

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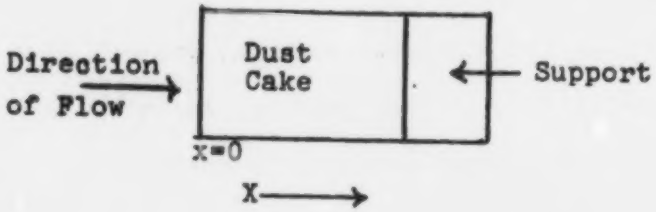
*The porosity could also be statistically evaluated with this data. The porosity is simply equal to the number of points selected lying in the open space divided by the total number of points.

DATA ANALYSIS

D'Arcy Flow

The following analysis has been found to be useful for evaluating pressure drop, velocity and time data obtained from studies on dust cakes collected on fibrous supports:



For laminar or creeping flow, D'Arcy's equation is applicable as follows: (see above figure)

$$\frac{dP}{dX} = -\frac{\mu V}{K_d} \qquad (25)$$

where P = pressure, dynes/cm^2

X = thickness, cm

μ = gas viscosity, poise

V = superficial velocity, cm/sec

K_d = D'Arcy permeability, cm^2

Equation (25) can be modified by substituting the following:

$$V = G/\rho \qquad \text{(Definition)} \quad (26)$$

$$\rho = \frac{PM}{RT} \qquad \text{(Perfect gas law)} \quad (27)$$

$$X = \frac{W}{\rho_t\ (1-\varepsilon)} \qquad \text{(Definition)} \quad (28)$$

$$K_d = K/\rho_t\ (1-\varepsilon) \qquad \text{(Definition)} \quad (29)$$

$$C = RT/M \qquad \text{(Definition)} \quad (30)$$

where G = mass velocity of gas, gm/(sec cm^2)

M = molecular weight of gas, gm/mole

R = universal gas constant, (dynes cm)/$^\circ$K mole)

T = absolute temperature, $^\circ$K

ρ = gas density, gm/cm^3

W = areal density of filter medium, gm/cm^2

ρ_t = true density of filter medium, gm/cm^3

ε = porosity of filter medium, dimensionless

K = permeability, gm/cm

C = gas property, cm^2/sec^2

Substituting equations (26) through (30) into (25) yields:

$$\frac{dP}{dW} = \frac{\mu CG}{PK} \tag{31}$$

The following assumptions are now made:

(1) The support "equilibrates" rapidly and hence the support remains unchanged for most of the period in which the cake is formed.*

(2) The cake is homogeneous in the direction of flow throughout the period in which the cake is formed.

Equation (31) can be integrated for any instant during the formation of a cake, yielding:

$$P^2_{1c} - P^2_{oc} = \frac{2\mu C\, GW_c}{K_c} \tag{32}$$

---

*In practice, "equilibrates" as used above simply means that the resistance of the support is constant.

Similarly, for the "equilibrated" support:

$$P^2_{1s} - P^2_{os} = \frac{2\mu C\ GW_s}{K_s} \qquad (33)$$

where the subscripts indicate:

i = inlet

o = outlet

c = particle cake

s = "equilibrated" support

Since the pressure on the outlet side of the cake (P_{oc}) is the same as the pressure on the inlet side of the support(P_{1s}), adding equations (32) and (33) yields:

$$P^2_{1c} - P^2_{os} = 2\mu CG\ (\frac{W_c}{K_c} + \frac{W_s}{K_s}) \qquad (34)$$

The areal density of the cake (W_c) depends on time as follows:

$$W_c = FGt \qquad (35)$$

Where F* = weight of particles/weight of gas, dimensionless

t = time required to form cake, sec

provided G and F are constant.

---

*This assumes that the weight of dust required to "equilibrate" the support is negligible compared to that of the cake, and that the mass collection efficiency is 100%. For more accuracy, F can be based on the actual weight of the particles in the cake.

Substituting equation (35) into (34) gives:

$$P^2_{ic} - P^2_{os} = \frac{2\mu CFG^2 t}{K_c} + \frac{2\mu CGW_s}{K_s} \qquad (36)$$

Therefore, if $(P^2_{ic} - P^2_{os})$* is plotted versus time, a straight line should be obtained whose slope and intercept are a measure of the permeability of the cake and "equilibrated" support respectively. For the initial brief period in which the filter support "equilibrates," this relationship with time will, of course, not necessarily follow a straight line. However, after this period, a linear relationship with time suggests the following:

(1) Compaction effects due to aerodynamic drag on the particles within the cake are not important. Hence, the porosity and permeability are homogeneous in the direction of flow.

(2) Those deposited particles which affect the cake permeability (all particles with the exception of possibly the very small ones) are collected at or close to the surface of the cake.

In other words, the assumptions made previously are valid. If cake compaction is important or if a significant number of particles are removed deep in the cake or in the support, a linear relationship with time will not be obtained unless compensations are made by other unknown effects. Rather, it would be expected that ($P^2_{1c} - P^2_{os}$) would rise more rapidly as time precedes.

---

* Note that $(P^2_{ic} - P^2_{os}) = 2\, P_{ave}\, \Delta P$

where P_{ave} = average of inlet and outlet pressure, dynes/cm^2

ΔP = pressure drop across cake and support, dynes/cm^2

Inertial Effects (Transition Flow)

Since the above discussion is based on D'Arcy's law as its starting point, it is only valid in the laminar flow regime. In order to demonstrate laminar flow, a velocity scan (pressure drop versus velocity) of the cake must be made. Using equation (34) if ($P^2_{1c} - P^2_{os}$) is plotted against the mass velocity then a straight line will be obtained if, and only if, the flow through both the cake and the support is laminar.

If a straight line is not obtained, inertial effects may be important. D'Arcy's equation takes the following form when corrected for inertial effects:

$$\frac{dP}{dX} = \frac{\mu V}{K_d} + \frac{V^2}{K'_d} \qquad (37)$$

Substituting equations (26) through (30) into (37) yields the following:

$$\frac{dP}{dW} = \frac{CG}{PK} + \frac{CG^2}{PK'} \qquad (38)$$

where K'_d = inertial constant, cm

K' = inertial permeability = $K'_d \, P_t \, (1-\varepsilon)$, gm/cm^2

For any instant in time during the formation of a cake equation (38) can be integrated for both the cake and the "equilibrated" support yielding:

$$P^2_{1c} - P^2_{os} = 2\mu CG \left(\frac{W_c}{K_c} + \frac{W_s}{K_s}\right) + 2CG^2 \left(\frac{W_c}{K'_c} + \frac{W_s}{K'_s}\right) \qquad (39)$$

Substituting equation (35) into (39) yields:

$$P^2_{1c} - P^2_{os} = 2CG^2F \left(\frac{\mu}{K_c} + \frac{G}{K'_c}\right)t + 2CGW_s \left(\frac{\mu}{K_s} + \frac{G}{K'_s}\right) \qquad (40)$$

As with the laminar flow case, if $(P^2_{1c} - P^2_{os})$ is plotted versus time a straight line should be obtained provided the assumptions made previously are valid (i.e. no compaction effects and most particles collected near cake surface). For transition flow, however, the slope and intercept from a plot of $(P^2_{1c} - P^2_{os})$ versus time are less useful yielding the following expressions:

$$\text{Slope} = 2CG^2F \left(\frac{\mu}{K_c} + \frac{G}{K'_c}\right) \qquad (41)$$

$$\text{Intercept} = 2CGW_s \left(\frac{\mu}{K_s} + \frac{G}{K'_s}\right) \qquad (42)$$

The unknowns are the permeametric constants K_c, K'_c, K_s, K'_s. Since there are only two equations and four unknowns, these constants cannot be evaluated. If a velocity scan is made, and $(P^2_{1c} - P^2_{os})$ is plotted versus the mass velocity (G), a polynomial curve is obtained of the form:

$$(P^2_{1c} - P^2_{os}) = aG + bG^2 \qquad (43)$$

The coefficients a and b can be evaluated by curve fitting. Comparing equations (15) and (19), the following expressions can be written:

$$a = 2\mu C \left(\frac{W_c}{K_c} + \frac{W_s}{K_s}\right) \qquad (44)$$

$$b = 2C \left(\frac{W_c}{K'_c} + \frac{W_s}{K'_s}\right) \qquad (45)$$

It might seem that since equations (41), (42), (44) and (45) contain four unknowns (K_c, K_s, K'_c, K'_s), they can be solved. Unfortunately, the equations are not independent and hence no unique solution is possible.

Hence, for the transition flow regime, it is necessary to make two velocity scans at different times during the cake's formation. When both velocity scans are curve fitted, the unknowns can be caluclated from the four coefficients using equations (43), (44), and (45).

Slip Flow

If the pressure of a gas flowing in a capillary is reduced until its mean free path is an appreciable fraction of the capillary radius, the flow rate exceeds that calculated from Poiseuille's law. This phenomenon is attributed to slippage of the gas at the walls of the capillary. Scheidegger(25) gives the following relationship for a similar effect on porous media:

$$-V = \frac{dP}{dx}\ \frac{\varepsilon^3}{\mu\ kS^2(1-\varepsilon)^2} + \frac{dP}{dx}\ \frac{1.74\varepsilon^2\lambda}{\mu\ kS(1-\varepsilon)} \qquad (46)$$

where k = Kozeny constant, dimensionless

S = total surface area of all particles/total volume of all particles, cm^{-1}

λ = mean free path of gas, cm

Substituting into equation (46), equations (26), (27), (28), (30) and the following:

$$K = \frac{\rho_t\ \varepsilon^3}{S^2k(1-\varepsilon)} \qquad (47)$$

$$\lambda = \frac{RT}{\sqrt{2\pi}\ N_o\ \sigma^2}\ \frac{1}{P} \qquad (48)$$

$$Z = \frac{1.74\ RT\ \rho_t\ \varepsilon^2}{\sqrt{2\pi}\ N_o\sigma^2\ kS} \qquad (49)$$

where N_o = Avogodro's number, molecules/mole

σ = diameter of gas molecule, cm

Z = slip flow coefficient, $(gm/sec\ cm)^2$

yields:

$$\frac{dP}{dW} = - \frac{\mu C G}{PK + Z} \quad (50)^*$$

Integrating from the inlet to outlet side of the cake yields:

$$\frac{P^2_{1c} - P^2_{oc}}{2} = \frac{\mu C W_c G}{K_c} - \frac{Z_c}{K_c} (P_{1c} - P_{oc}) \quad (51)^{**}$$

If we assume that the support is "saturated" (i.e. the resistance of the support does not depend on when during the cake's formation the velocity scan was made), the pressure at the interface of the cake and the support depends only on the mass velocity of the gas and the pressure on the downstream side of the support. Hence, if velocity scans are done at two different times, P_{oc} (the pressure at the interface) will be constant at a given flow rate if the pressure on the downstream side of the support is constant. Writing equation (51) for time t_1 and t_2 yields:

$$\frac{P^2_{1c(at\ t_1)} - P^2_{oc}}{2} = \frac{\mu C W_{c(at\ t_1)}}{K_c} - \frac{Z_c}{K_c} (P_{1c(at\ t_1)} - P_{oc}) \quad (52)$$

---

*This reduces to equation (31) if slip is unimportant.

** This reduces to equation (32) if slip is unimportant.

and

$$\frac{P^2_{1c(at\,t_2)} - P^2_{oc}}{2} = \frac{\mu C W_{c(at\,t_2)}\, G}{K_c} - \frac{Z_c}{K_c}(P_{1c(at\,t_2)} - P_{oc}) \qquad (53)$$

Subtracting equation (52) from equation (53) yields:

$$\frac{P^2_{1c(at\,t_2)} - P^2_{1c(at\,t_1)}}{2} = \frac{\mu C(W_{c(at\,t_2)} - W_{c(at\,t_1)})G}{K_c} - \frac{Z_c}{K_c}(P_{1c(at\,t_2)} - P_{1c(at\,t_1)}) \qquad (54)$$

Rearranging gives:

$$P_{ave} = \frac{\mu C\Delta W}{K_c}\,\frac{G}{P} - \frac{Z_c}{K_c} \qquad (55)$$

where:

$$P_{ave} = \frac{P_{1c(at\,t_2)} + P_{1c(at\,t_1)}}{2}$$

$$\Delta P = P_{1c(at\,t_2)} - P_{1c(at\,t_1)}$$

$$\Delta W = W_{c(at\,t_2)} - W_{c(at\,t_1)}$$

Plotting P_{ave} versus $(G/\Delta P)$ should give a straight line with the slope equal to $(\mu C\Delta W/K_c)$ and the intercept equal to (Z_c/K_c).

The above analysis is valid only if the downstream pressure is held constant. (The flow rate is controlled by adjusting the upstream pressure). Although the above analysis is correct, it is not wise to calculate coefficients quantifying slip from experimental data unless the pressure drop is significantly affected by slip. For this reason and in order to make the solution more tractable, it may be necessary to make velocity scans at very low flow rates where pressures are relatively low and compressibility unimportant. Rearranging equation (46) we obtain:

$$-V = \frac{dP}{dx}\frac{\varepsilon^3}{\mu k S^2(1-\varepsilon)^2} \quad (1 + \frac{1.74\lambda S(1-\varepsilon)}{\varepsilon}) \tag{56}$$

Letting

$$Q = 1 + \frac{1.74\lambda S(1-\varepsilon)}{\varepsilon} \tag{57}$$*

and substituting equations (28), (47), and (56) into (57) yields:

$$\frac{dP}{dW} = - \frac{V}{KQ} \tag{58}$$

Integrating equation (58) from the inlet to the outlet side of the cake gives:

$$P_{oc} - P_{1c} = - \frac{\mu V}{K_c Q_c} W_c \tag{59}$$

Writing an analogous equation for the support and adding it to equation (59) gives:

$$\Delta P = P_{1c} - P_{os} = \mu V \left(\frac{W_s}{K_s Q_s} + \frac{W_c}{K_c Q_c}\right) \tag{60}$$

If at different times during the formation of the cake two velocity scans are made at the same total pressure, values for

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*This is very similar in form to the Cunningham correction factor for particles.

K_sQ_s and K_cQ_c can be obtained. Now, if we assume that:

$$Q = 1 + \frac{A}{P} \qquad (61)$$

where A = a constant

which follows from equation (57), then multiplying equation (61) by K yields:

$$KQ = KA \frac{1}{P} + K \qquad (62)$$

A plot of KQ versus $(\frac{1}{P})$ for either the cake or the support will yield values for the permeability, K, and the constant, A.

Compaction Effects

If compaction effects are suspected, (If $(P^2_{1c} - P^2_{os})$ versus time has a positive second derivative) further velocity scans at still different times during the cake's formation can be made to quantitate the change in the permeametric constants due to compaction.

Compaction effects in powder beds are often described by exponential expressions, such as the following:(14)

$$\varepsilon = \varepsilon_o \exp(-aF) \qquad (63)$$

Where ε = porosity

ε_o = initial porosity

a = constant, cm^2/dynes

F = compressive stress, dynes/cm^2

Differentiating equation (63) yields:

$$\frac{d\varepsilon}{dF} = - a\varepsilon \qquad (64)$$

Differentiating equation (28) gives:

$$dx = \frac{dW}{(1-\varepsilon)\rho_t} + \frac{W}{\rho_t} \frac{d\varepsilon}{(1-\varepsilon)^2} \qquad (65)$$

Combining equations (25) and (65) yields:

$$- \frac{K_d}{\mu V} dP = \frac{dW}{(1-\varepsilon)\rho_t} + \frac{W}{\rho_t} \frac{d\varepsilon}{(1-\varepsilon)^2} \qquad (66)$$

from the Kozeny-Carman Equation:

$$K_d = \frac{\varepsilon^3}{kS^2(1-\varepsilon)^2} \qquad (67)$$

Combining equations (66) and (67) yields:

$$dP = - \frac{\mu k S^2 V}{\rho_t} \left(\frac{(1-\varepsilon)}{\varepsilon^3} dW + \frac{W}{\varepsilon^3} d\varepsilon \right) \qquad (68)$$

For a small increment of cake:

$$dF = -dP \qquad (69)$$

Combining equations (64), (68), and (69) yields:

$$\frac{dW}{d\varepsilon} = \frac{A \ \varepsilon^2 + W}{(\varepsilon - 1)} \qquad (70)$$

where A = constant, gm/cm^2 = $\dfrac{\rho_t}{a\mu VkS^2}$

Solving equation (70) with boundary conditions $\varepsilon = \varepsilon_o$ at $W = 0$ yields:

$$\frac{W}{A} = \varepsilon^2 - 2\varepsilon - 2(1-\varepsilon)\ln(1-\varepsilon) - (1-\varepsilon)\left(\frac{2\varepsilon_o - \varepsilon_o^{\,2} + 2(1-\varepsilon_o)\ln(1-\varepsilon_o)}{(1-\varepsilon_o)}\right) \quad (71)$$

Since the total compressive stress on the cake is equal to P, equation (63) can be rewritten as:

$$\varepsilon = \varepsilon_o \exp(-a\Delta P) \quad (72)$$

Combining equations (71) and (72) yields:

$$\frac{W}{A} = \varepsilon_o^{\,2}[\exp(-a\Delta P)]^2 - 2\varepsilon_o\exp(-a\Delta P) - 2[1-\varepsilon_o\exp(-a\Delta P)]\ln[1-\varepsilon_o\exp(-a\Delta P)]$$
$$- [1-\varepsilon_o\exp(-a\Delta P)]\left[\frac{2\varepsilon_o-\varepsilon_o^{\,2} + 2(1-\varepsilon_o)\ln(1-\varepsilon_o)}{(1-\varepsilon_o)}\right] \quad (73)$$

combining equation (35) and (73) yields:

$$\frac{t}{T} = \varepsilon_o^{\,2}[\exp(-a\Delta P)]^2 - 2\varepsilon_o\exp(-a\Delta P)] - 2[1-\varepsilon_o\exp(-a\Delta P)]\ln[1-\varepsilon_o\exp(-a\Delta P)]$$
$$- [1-\varepsilon_o\exp(-a\Delta P)]\left[\frac{2\varepsilon_o-\varepsilon_o^{\,2} + 2(1-\varepsilon_o)\ln(1-\varepsilon_o)]}{(1-\varepsilon_o)}\right. \quad (74)$$

where T = constant, sec = ${}^{\rho}t^{\rho}/FG^{2}a\mu kS^{2}$

A plot of pressure versus time for cake-build-up can be curve fitted to equation (74) and the constants a, ε_o, and T evaluated by regression analysis.

EXPERIMENTAL RESULTS

The pressure drop was recorded as a function of time (and hence areal density since the mass flow rate of dust is constant). At the end of a run, and sometimes during a run, the dust flow was halted, and a velocity scan (i.e. determination of pressure drop as a function of gas flow rate) was made. The velocity, however, was never allowed to exceed the value at which the cake was formed. The weight of the dust cake, its diameter, and thickness were then measured.

Experimental Problems

Three criticisms regarding the equipment design for runs 3,4, and 7 can be made. (1) As the cake builds up the deposition velocity decreases.* Hence, it might be expected that the particles would pack less tightly as cake thickness increased. (2) The particle size distribution varied with flow rate since no elutriator was used. Therefore, comparison of runs at high and low flow rate may be misleading because the aerosol also changed. (3) No charge neutralizer was employed.

For runs 9, 11, 12 and 14 the same criticisms hold as for runs 3, 4, and 7 except that the approach velocity was held constant. Also upstream pressures of at least 95 psia were required to drive the gas through the dust feeder, the saturated paper, and a reasonable thickness of cake.

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*The mass flow of air was held constant by increasing upstream pressure as the pressure drop across the cake increased.

Pressure Drop Versus Time

Referring to Figures 17 and 18 for runs 3 and 9 respectively, it can be seen that a plot of pressure square difference* versus deposition time is a straight line except for the initial period of cake build-up. These plots are representative of all runs mentioned above. The linearity suggests that permeability, porosity, and pore size distribution are homogeneous in the direction of flow. It further implies that compaction effects due to aerodynamic drag on the cake were not significant. If the above statements are untrue the $\Delta(P^2)$ versus time relationship would bend.**

Referring again to Figures 17 and 18 it is seen that the initial segment of the curve is concave downward for run 3 and concave upward for run 9. Since run 3 used Whatman 44 paper as a support, whereas run 9 used S & S 589-1H, it seems likely that this was responsible for the difference. Whatman 44 is a very retentive paper that would be expected to plug rapidly due to small particles trapped in its pores. S & S 598-1H, on the other hand, is a very open paper. Small particles would initially pass through the paper and the larger particles which would be retained would probably leave a structure composed of large pores as the paper saturates.

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*$\Delta(P^2)$ = (Pressure at inlet of cake)^2 - (pressure at outlet of support)^2 = $P^2_{ic} - P^2_{os}$

**The effects of cake compaction may also be offset by the variable deposition velocity in runs 3, 4, and 7.

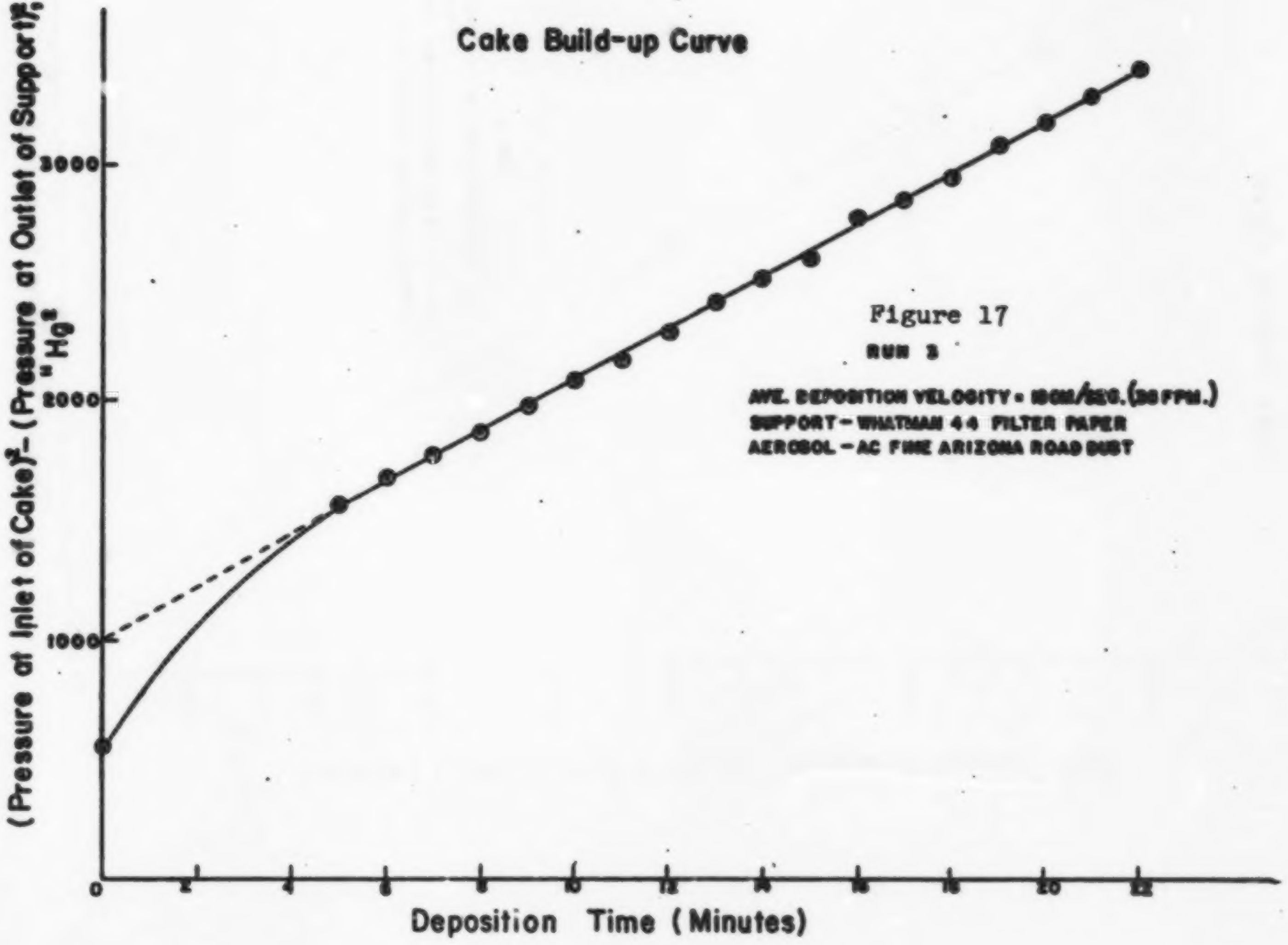

Figure 17

RUN 3

AVE. DEPOSITION VELOCITY = 10CM/SEC. (20 FPM.)
SUPPORT - WHATMAN 44 FILTER PAPER
AEROSOL - AC FINE ARIZONA ROAD DUST

Cake Build-Up Curve

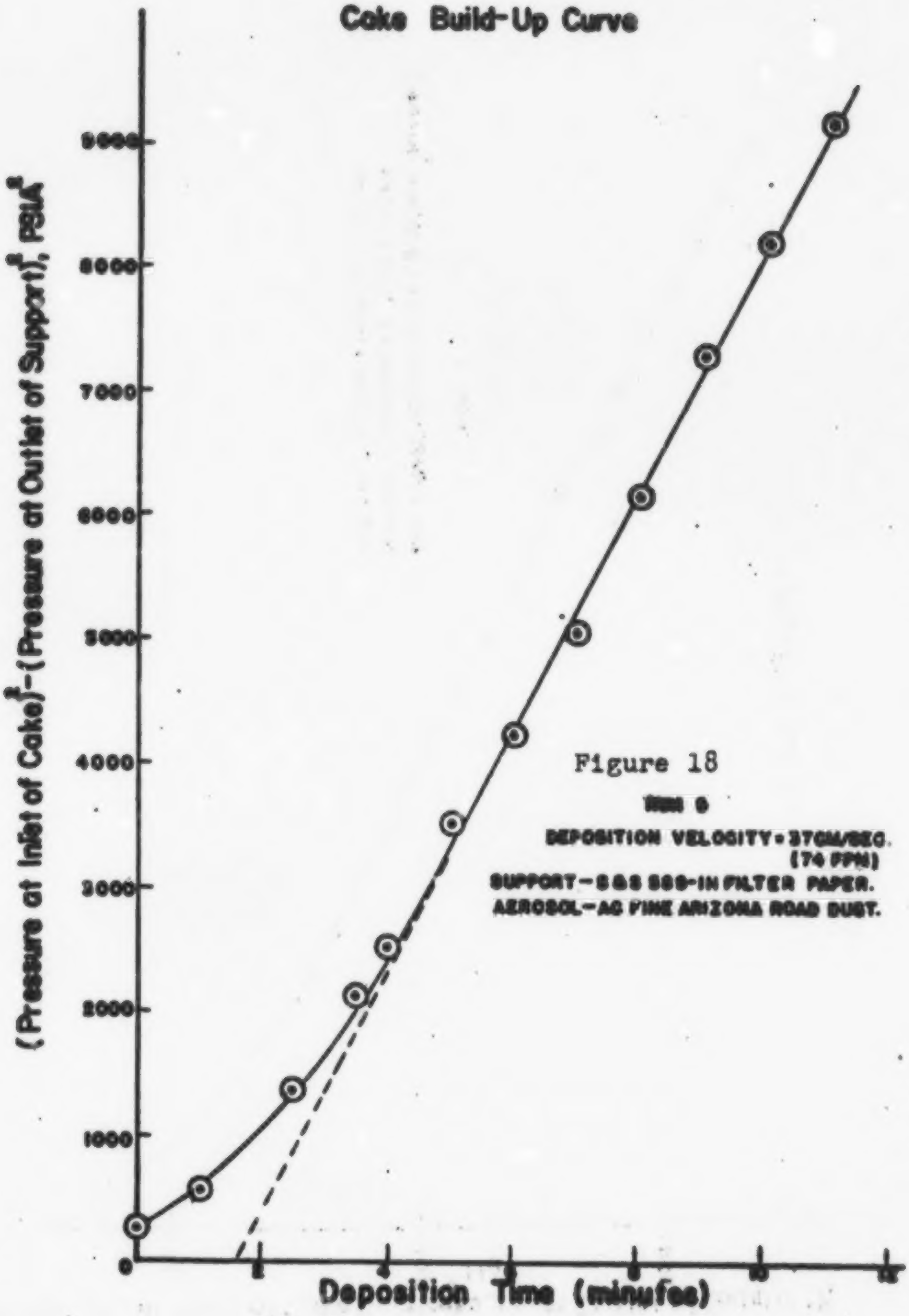

Figure 18

RUN 6

DEPOSITION VELOCITY = 37CM/SEC.
(74 FPM)
SUPPORT - S&S 589-IN FILTER PAPER.
AEROSOL - AC FINE ARIZONA ROAD DUST.

Velocity Scans

A velocity scan of the cake made in run 9 showed an interesting effect. When the pressure square difference was plotted as a function of mass velocity of the gas, it was found that as the velocity increased, the pressure square difference increased less than a linear relationship would predict. Runs 11 and 12 were made to duplicate run 9 and verify the effect noted above. (See Figure 19 for a typical velocity scan.) This effect was opposite to that seen in runs 3, 4, and 7 (See Figure 20.) In these runs the pressure square difference increased more rapidly with the mass velocity of the gas than a linear relationship would predict. This was originally attributed to inertial effects although the Ergun* equation predicted that inertial effects should be insignificant.

In runs 9, 11, 12 and 14 a velocity scan was made by holding the upstream pressure constant, and decreasing the downstream pressure to increase flow rate. If a condition of D'Arcy flow (creeping flow) exists it would be expected that the pressure square difference is a linear function of mass flow rate.

However, it is well known (25) that if the mean free path of a gas approaches the pore size of the porous media, there will be a slip at the walls of the pores (in other words, the gas very close to the walls of a pore cannot be assumed stationary as in

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* See Reference (10).

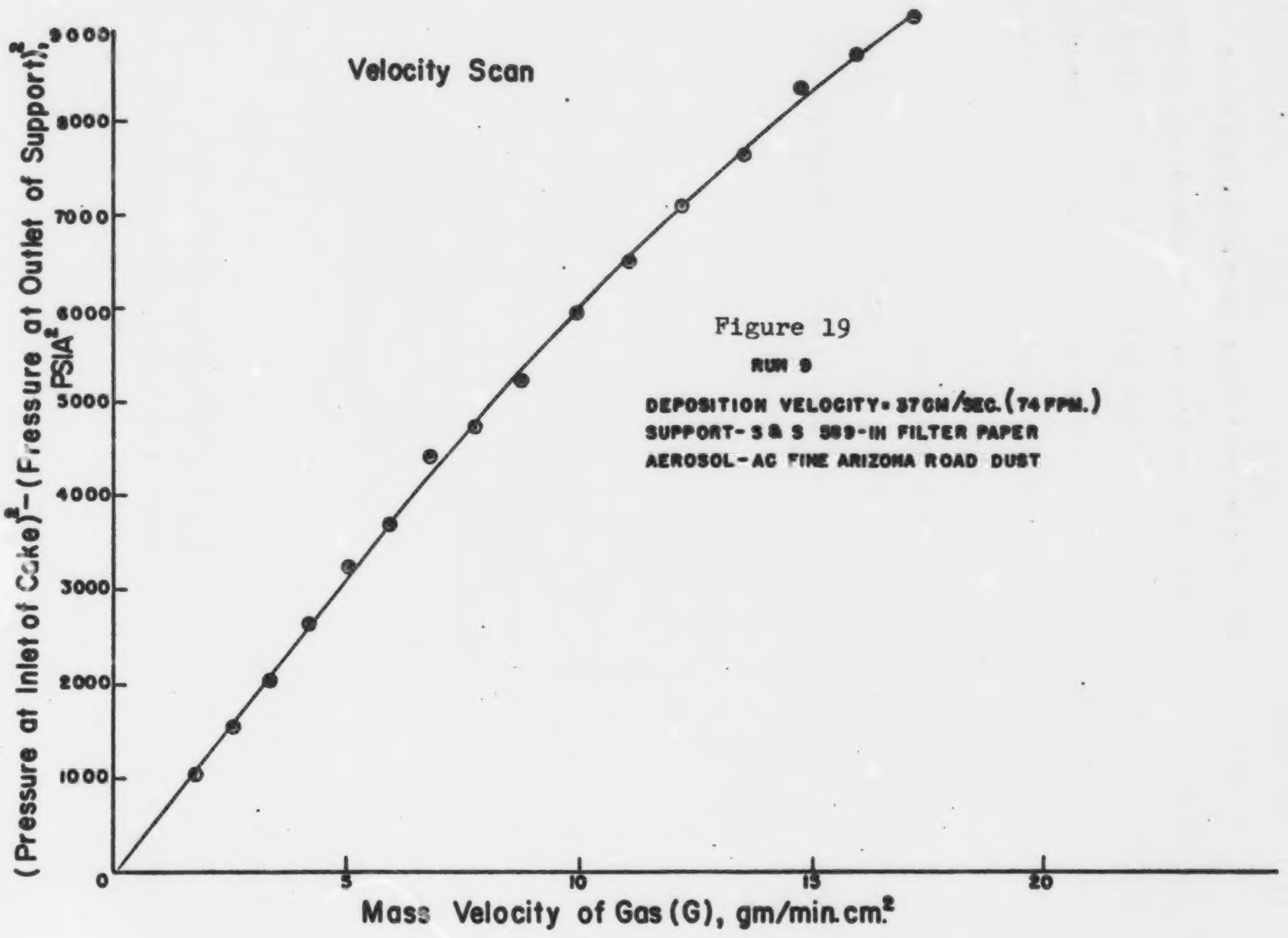

Figure 19

RUN 9

DEPOSITION VELOCITY = 37 CM/SEC. (74 FPM.)
SUPPORT - S & S 589-IH FILTER PAPER
AEROSOL - AC FINE ARIZONA ROAD DUST

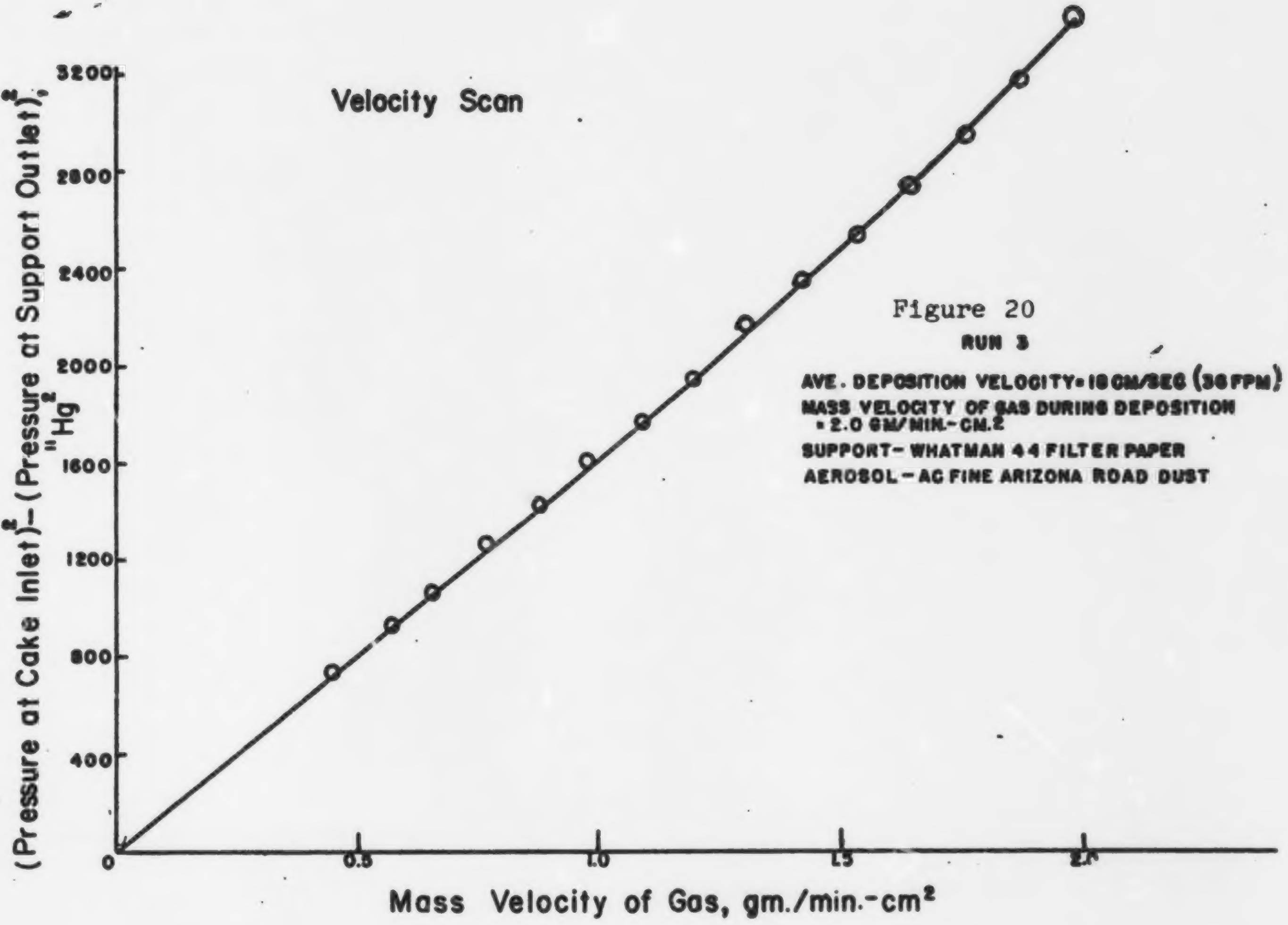

Figure 20

RUN 3

AVE. DEPOSITION VELOCITY = 18 CM/SEC (36 FPM)
MASS VELOCITY OF GAS DURING DEPOSITION
= 2.0 GM/MIN.-CM.2
SUPPORT - WHATMAN 44 FILTER PAPER
AEROSOL - AC FINE ARIZONA ROAD DUST

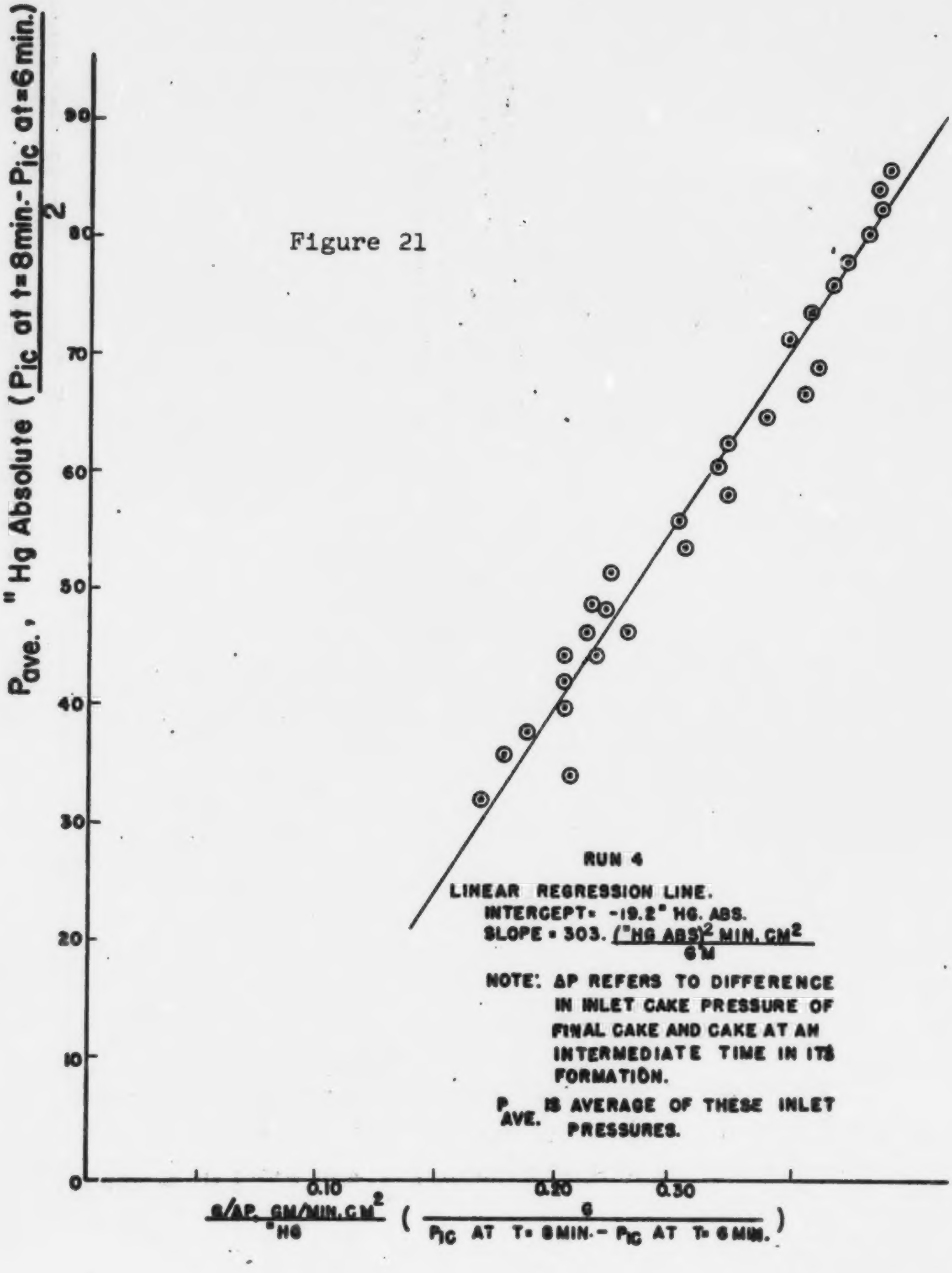

Figure 21

P_ave., "Hg Absolute ((P_IC at t=8min.-P_IC at=6min.) / 2)
0
10
20
30
40
50
60
70
80
90
0.10
0.20
0.30
G/ΔP, GM/MIN.CM² / "HG (G / P_IC AT T= 8MIN.- P_IC AT T= 6MIN.)
RUN 4
LINEAR REGRESSION LINE.
INTERCEPT= -19.2" HG. ABS.
SLOPE = 303. ("HG ABS)² MIN. CM² / GM
NOTE: ΔP REFERS TO DIFFERENCE IN INLET CAKE PRESSURE OF FINAL CAKE AND CAKE AT AN INTERMEDIATE TIME IN ITS FORMATION.
P_AVE. IS AVERAGE OF THESE INLET PRESSURES.

the case for fully viscous flow). This phenomenon has the effect of decreasing the pressure drop (or pressure square difference) at a given velocity from that predicted by empirical or theoretical relationships based on no slip.

Since in runs 9, 11, 12, and 14, the absolute pressure was relatively high for the velocity scans in the lower flow range, the mean free path of the gas is small. If at these conditions, the phenomena of slip is relatively unimportant and the flow rate is increased by decreasing the pressure downstream of the filter cake about 5-fold (thereby increasing the mean free path by the same amount), slip flow in the pores near the exit of the cake may be possible. Hence, a smaller pressure square difference would be obtained than expected.

In the runs 3, 4, and 7, a velocity scan was made by increasing the upstream pressure to increase flow rate while the downstream prsssure was held constant at one atmosphere. Hence in the low flow rate range, the absolute pressure throughout the cake was near atmospheric pressure and slip may have been important. This would yield a smaller pressure square difference for a given mass velocity of the gas than would be expected. As the flow rate is increased, the pressure upstream of the filter cake increases significantly. Slip flow may then be relatively unimportant in the upstream portion of the cake and hence the pressure square difference would be higher for a given flow rate than with slip. This would yield a curve in which the pressure square difference increases more rapidly with flow rate than a linear relationship.

For run 4 a plot of $\Delta(P^2)$ versus flow rate yielded a non-linear curve. As described in equation 55, a plot of P_{ave} versus G/P_{1c} should yield a straight line if there exists a condition of slip flow. Figure 18, a plot of those parameters, suggests that slip can explain the curvature of the velocity scans. A similar plot for other runs was not possible.

Porosity and Permeability

Values for porosity and permeability for those runs least troubled with operating problems are given in Table 7.

Table 7

Cake Porosity and Permeability Data

Run No.	Support Paper	Deposition Velocity cm/sec (fpm)	Cake Porosity	Permeability (gm/cm) x 10^{11}
3	Whatman 44	18 (36)*	0.63	9.4
4	Whatman 44	18 (36)*	0.63	9.7
7	Whatman 44	21 (43)	0.80	2.2
9	S & S 589-1H	37 (74)	0.73	5.5
11	S & S 589-1H	37 (74)	0.71	5.0
12	S & S 589-1H	37 (74)	0.69	4.3
14	Whatman 44	37 (74)	0.77	4.1

Figure 22 indicates a strong, inverse relationship between permeability and deposition velocity (approach velocity at which cake formed). This can probably be attributed to the increased momentum of the particles at the higher approach velocity as they

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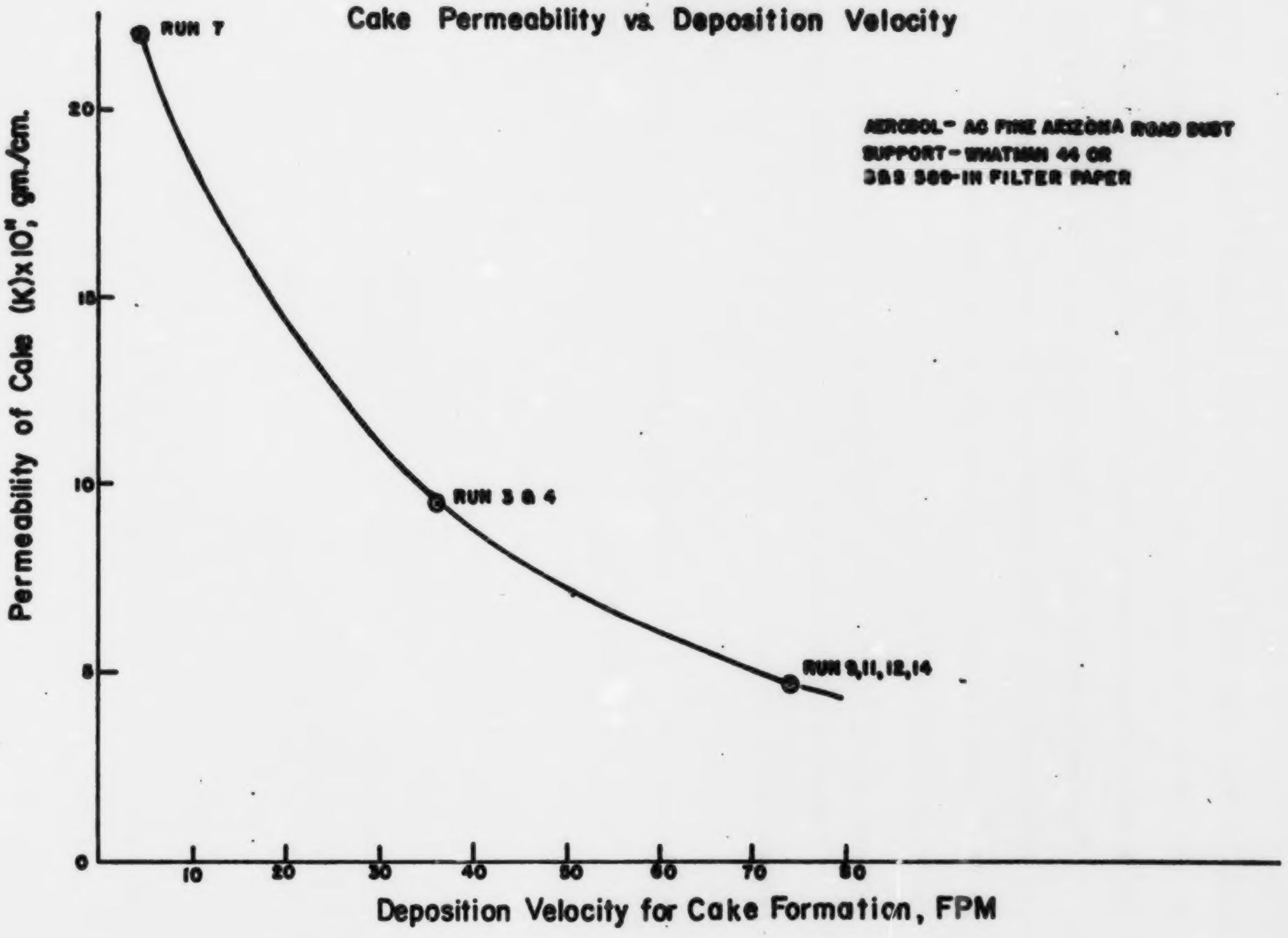
*Deposition velocity was 45 fpm at beginning of run and 27 fpm at end of run.

Figure 22

Cake Permeability vs. Deposition Velocity



land on the cake and because of disruption of particle "trees" at the surface of the cake. Both of these effects would tend to leave more compact structures at the surface of the cake.

The porosity of the filter cakes does not correlate with the permeability. This is probably, in part, due to elutriation of the inlet aerosol because of different approach velocities and different flow geometries.

Comparison of Results with Literature

Experimental studies on cake filtration reported in the literature have been extremely meager. In one of the more prominent studies, Williams, et al(19) obtained values for the permeability, K, for cakes of various compositions and particle sizes. The range of values obtained are tabulated below along with data of this experimenter for comparison:

Table 8

Cake Permeability Data

Source	Approach Velocity cm/sec	Particle Size, μm	Viscous Permeability $K_c \times 10^{11}$, gm/cm
Williams, et al	5	20	38 - 110
Run 7	2	1.9**	22
Run 3 & 4	16	1.9**	10
Run 9, 11, 12 & 14	37	1.9**	4.7

The permeability of Williams, et al is somewhat higher but not significantly considering that they evaluated particle size only qualitatively.

**Sauter mean diameter.

Billings and Wilder(26) reviewed a large number of reference sources and conducted a questionnaire-interview survey amongst fabric filter users to obtain values of the permeability, K. Although, data on conditions of operations were limited and results contradictory, they devised a correlation to predict K. Based on their correlation, results obtained are of the same order of magnitude as those obtained by this experimenter. More exact comparisons are not possible due to the crudeness of the correlation. They also derived an inverse relationship between K and approach velocity ($K \propto \frac{1}{V}$). This is in agreement with two of the three points plotted in Figure 22.

Models devised to predict permeabilities for cake filtration have relied mainly on those available for packed beds. The model of Kozeny-Carman for laminar flow is the one usually chosen. This would yields the following expression for the permeability, K:

$$K = \frac{D^2_{sv} \rho_t \varepsilon^3}{36k\ (1-\varepsilon)} \tag{75}$$

where K = permeability, gm/cm

ε = porosity

ρ_t = density of the dust, gm/cm^3

D_{sv} = Sauter mean diameter (total surface area of all particles/total volume of all particles), cm

k = Kozeny constant = $2\tau^2 \approx 4.2$ (from large amount of experiment data)

τ = Tortuosity (average distance traveled passing through bed/actual bed depth)

Calculating the permeabilities from the measured values of particle size distribution, porosity, and true particle density yields the following:

Table 9

Predicted and Experimental Permeabilities

Run No.	Deposition Velocity (cm/sec)	Permeability, $K_c \times 10^{11}$ (gm/cm) Experimental	Predicted
3	18	9.4	47
4	18	9.7	47
7	2.1	22	180
9	37	5.5	100
11	37	5.0	86
12	37	4.3	74
14	37	4.1	140

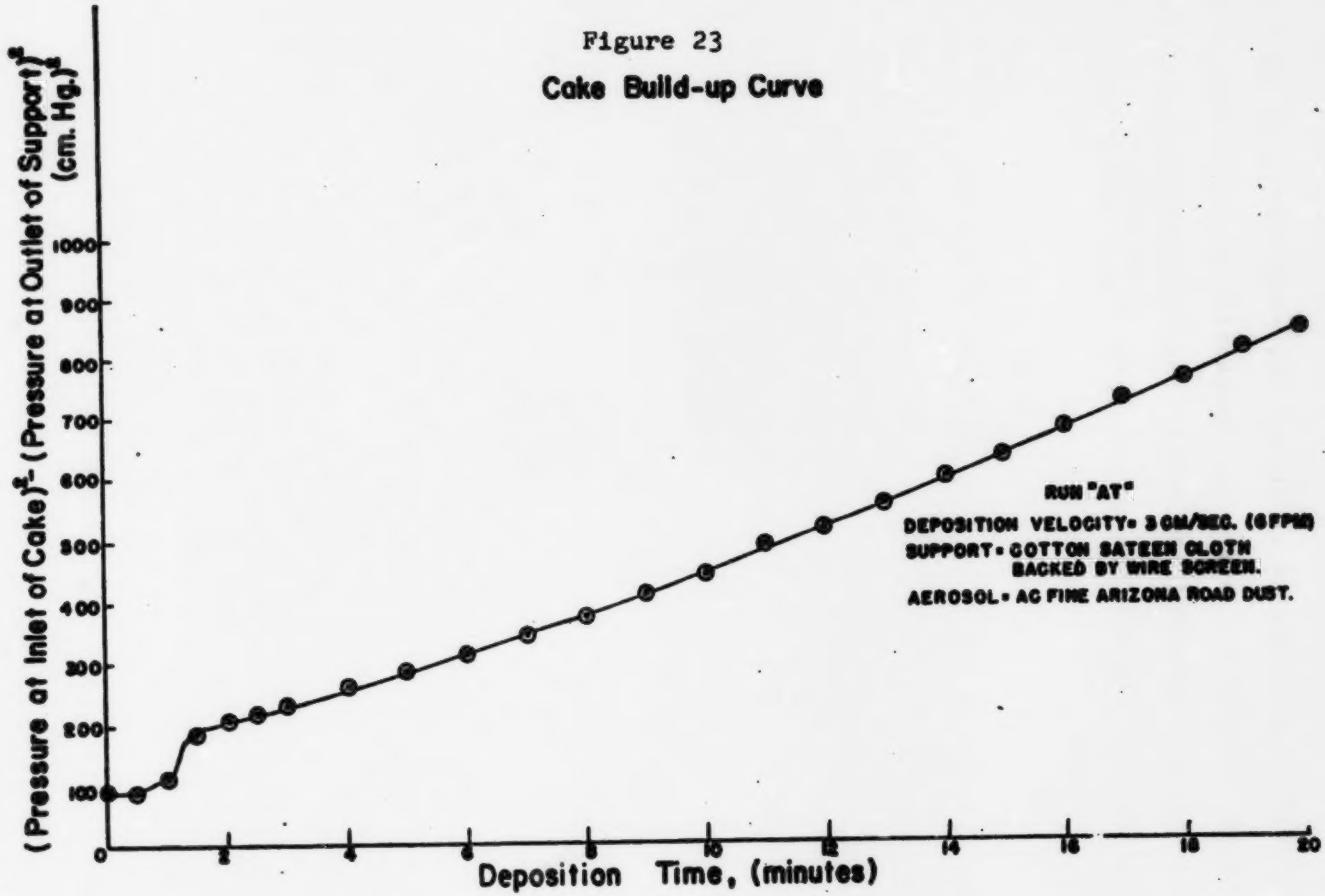
The predicted values for permeability are always higher than the experimental values. The experimental values of permeability show a reasonable trend with deposition velocity while the predicted values show no trend at all.

Results from Latter Runs

The latter runs (not yet discussed) were conducted at 6 fpm deposition velocity using experimental procedures discussed above. The runs were repeated many times, and each run gave a curved (concave upward) $\Delta(P^2)$ versus deposition time line for cake build-up. (see Figure 23 for example). Previously straight lines had been obtained. The reason for this discrepancy is not understood, but it is believed that the newer results are probably correct because the experimental design and operation were more precise than in previous work. Obviously, further work

Figure 23

Cake Build-up Curve



would be required to be sure. The most likely explanations for this nonlinearity are: (1) Compaction and (2) Change in cake or support permeability due to the filtering of particles not removed at the surface of the cake.

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Section VI

APPLICABILITY OF CONVENTIONAL BAGHOUSES TO HIGH VELOCITY OPERATION

GENERAL

The objective of this study is to evaluate the feasibility of higher air to cloth ratios thanare commonly employed in fabric filtration. Aside from this, the study should also provide a base against which potential improvements can be judged. Pressure drop and mass collection efficiencies were determined for two pilot plant size bag houses. The first unit tested was a pulse-jet cleaned filter assembled mostly from parts obtained from Buffalo Forge Company and used in their marketed units. An attempt to operate the unit at typical conditions was not successful with respect to both pressure drop stability and collection efficiency. The reasons for this failure are not entirely known. It may be due to a peculiar interaction of fabric (polyester felt) and dust (fly ash).

The second unit tested was a shaker cleaned filter previously employed for cleaning gases from a sodium fire. The unit was used essentially without modification (including woven teflon bag) at an air to cloth ratio of 10:1. (During the sodium filtration work it was operated at 2:1 air to cloth ratio.) Although removal efficiencies were adequate (-99.95 wt %), the collector would not equilibrate with respect to pressure drop for repeated filtration cycles. This instability was probably due to the design of the dust hopper.

Both the pulse-jet and shaker filter were operated for brief periods at higher velocities (up to 50 CFM/FT^2). The problems noted above were significantly worse than at 10 CFM/FT^2 air to cloth ratio.

PULSE-JET EXPERIMENTS

The pulse-jet filter used in this experimental program was a collection of parts from a commercial unit* mounted in a specially designed pilot plant size housing (See Figure 24). Two polyester bags (16oz/yd^2, approx. 15 μm fiber diameter), 8 feet long x 4 1/2 inches diameter were hung vertically in a 13 1/4 inch diameter cylindrical housing. Each bag was supported on a cylindrically-shaped wire cage which was attached to the underside of a clean air plenum located at the top of the unit. The metal cage prevented bag collapse, since in normal operation dust-laden gas entered the dirty air plenum surrounding the bags (depositing dust on the outer surface) and filtered gas passed through the bags to the clean air plenum. The inlet was placed near the base, somewhat above the dust hopper located below the bags.

Bag cleaning was accomplished in the following manner. A pulse of air was ejected from a 1/4 inch I.D. nozzle located above and directed into the mouth of a venturi section attached to the top of each bag. The nozzle tip was located about 3 inches from the venturi inlet. The degree of cleaning could be controlled by adjusting the cleaning air pressure, the duration of the cleaning pulse or the rate of pulsing.

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*obtained from Buffalo Forge Company

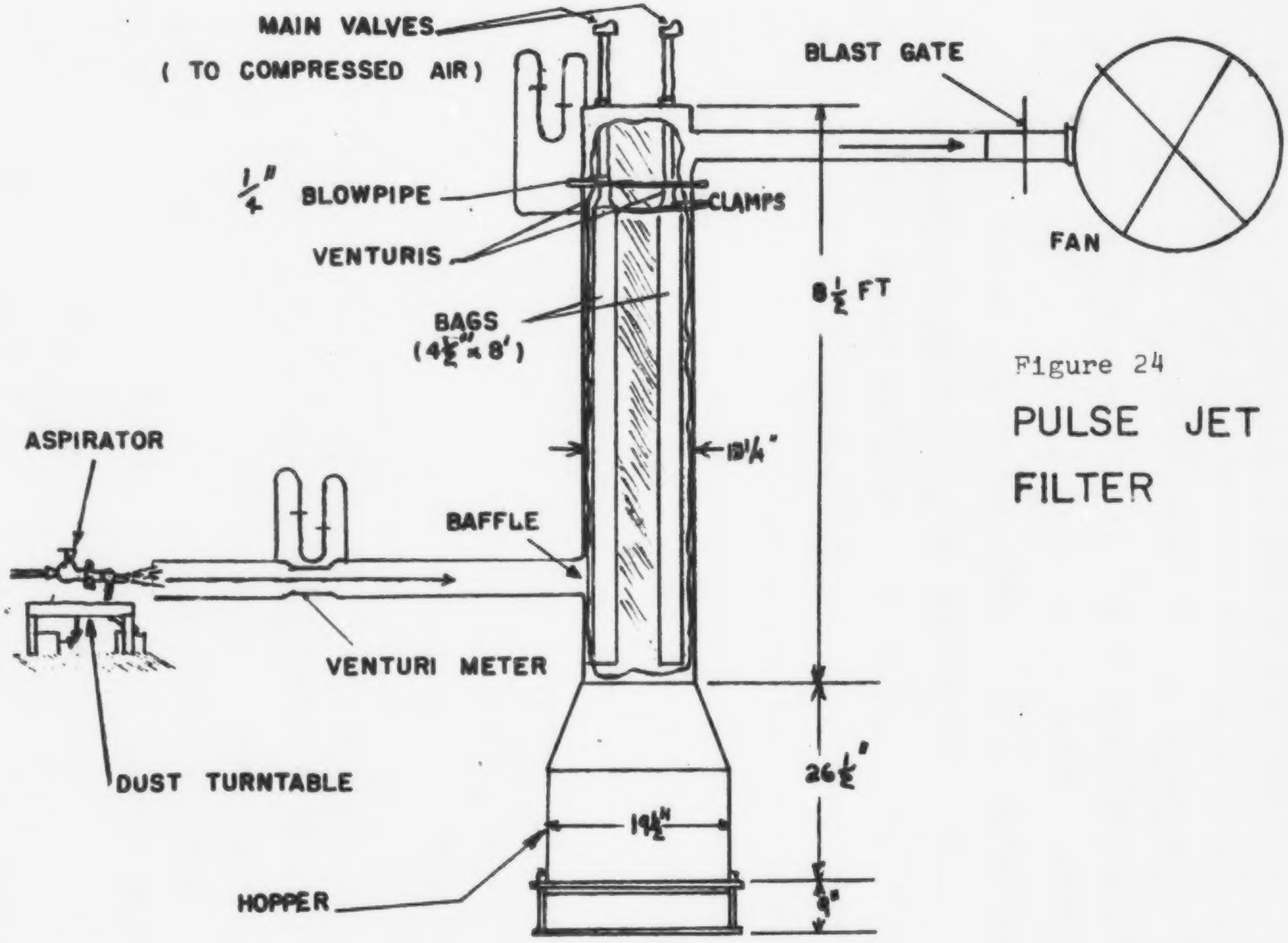
MAIN VALVES
(TO COMPRESSED AIR)
BLAST GATE
$\frac{1}{4}''$ BLOWPIPE
CLAMPS
VENTURIS
FAN
$8\frac{1}{2}$ FT
BAGS
($4\frac{1}{2}'' \times 8'$)
ASPIRATOR
$13\frac{1}{4}''$
BAFFLE
VENTURI METER
DUST TURNTABLE
$26\frac{1}{2}''$
$19\frac{1}{2}''$
HOPPER
9"

Figure 24
PULSE JET
FILTER

Precipitator fly ash (mean size by count = 0.6 μm ; $\sigma g=3.1$) was injected into the collector inlet duct by a compressed-air ejector operated at 35 psig to obtain good particle dispersion. Dust to the ejector suction was regulated by the Harvard turntable generator described by Silverman and Billings[1]. Inlet dust loading was about 1 grain per cubic foot air.

Inlet dust samples were collected isokinetically on Whatman paper thimbes or on MSA 1106B glass filter discs. Outlet dust samples were collected on MSA 1106B filter discs. Loadings were determined gravimetrically.

Gas flow rates were measured by a venturi flow meter located upstream of the collector which had been previously calibrated by a pitot tube traverse. Collector resistance measurements were based on the pressure differential between the gas inlet chamber and exit plenum. A Buffalo Forge high static pressure fan and a damper was used downstream of the collector.

Prior to the start of this experimental study, we were not able to equilibrate the pulse-jet cleaned filter at constant pressure drop. Repetition of this work verified this instability. The operating conditions were as follows:

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[1] Silverman, L and Billings, C.E., "Methods of Generating of Solid Aerosols", JAPCA, 6:76 (1956)

Inlet dust loading - approx. 1_2gr/ft^3
Air/cloth ratio - 10 CFM/FT^2

Cleaning air pulse duration - 500 msec
Pulse rate - 4 pulses per bag per min*
Cleaning air reservoir pressure - 110 psig

The inlet dust loading and air to cloth ratio are typical for pulse-jet units (higher inlet concentrations are common). The cleaning cycle, if anything,was more severe than is normally encountered in this type of unit.

At the time it was believed that to stabilize the pressure drop, the severity of cleaning would have to be increased. This was done and a stable pressure drop of 7 to 8" H_2O was obtained for a 4 1/2 period as shown in Figure 25. The cleaning was as follows: Each bag was pulsed three times every 15 seconds alternately (so that neither bag was pulsed simultaneously). Each of the three pulses lasted for 400 msec with 250 msec dividing the three pulses. Cleaning air pressure was maintained at 110 psig.

It was discovered that the above described cleaning cycle yielded very poor mass efficiencies. They averaged 92 wt% (78.4 to 96.9% for 11 samples). Between these samples the following changes were made:

(1) Bags were removed and examined for tears, leaks at seams and at seals (none found)

(2) Seals between bags and venturis were modified to insure that there were no leaks

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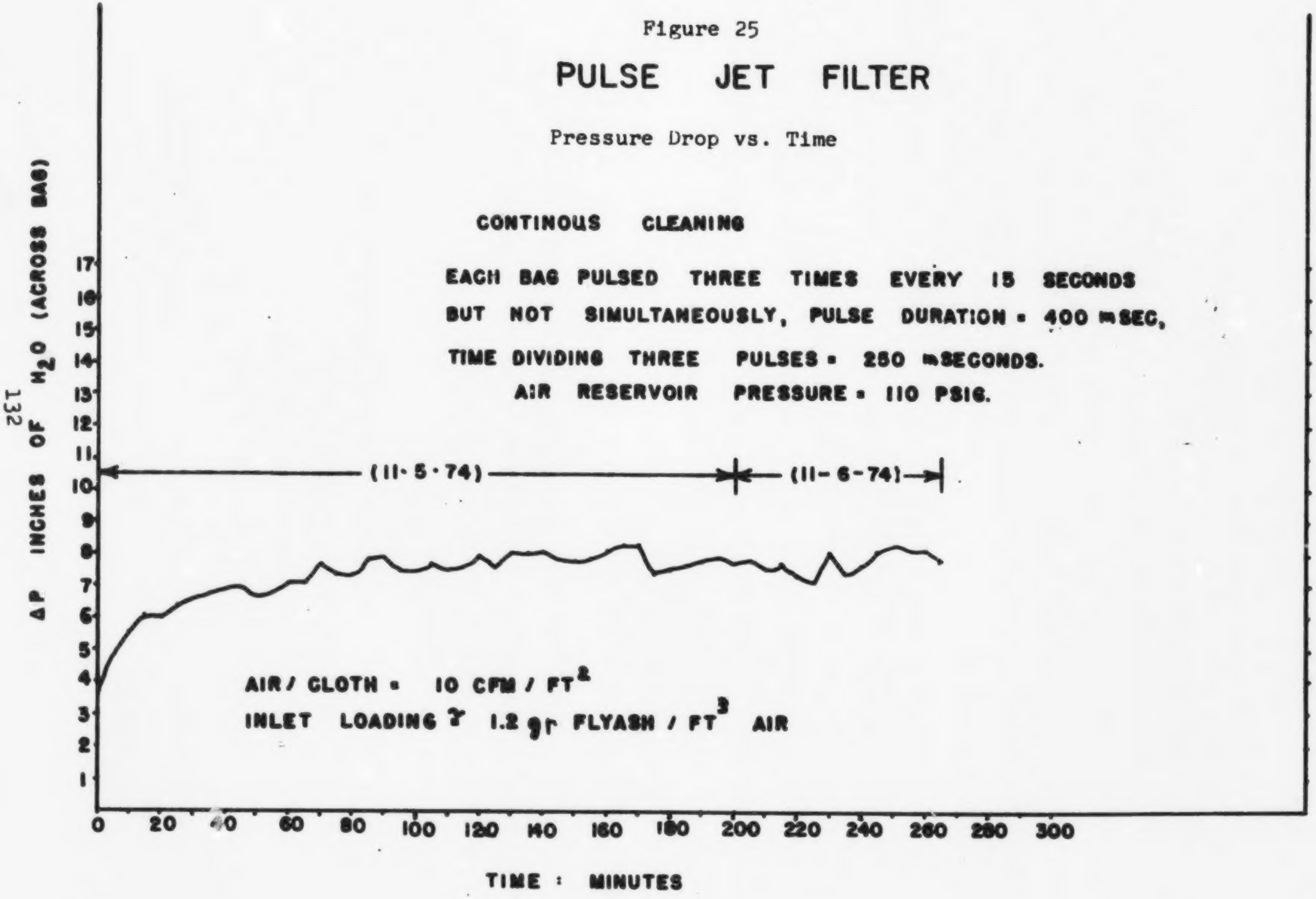
*Bags were not pulsed simultaneously. Each bag pulsed every other 7 1/2 seconds.

Figure 25

PULSE JET FILTER

Pressure Drop vs. Time



(3) New bags (duplicates of the original ones) were installed

(4) Compressed air reservoir pressure reduced from 110 psig to 90 psig

(5) Each bag was pulsed once rather than thrice every 15 seconds

(6) Pulse duration (electrical time) was reduced from 400 msec to 45 msec

The higher efficiencies noted were obtained after the changes were made. Measuring the outlet dust loadings with a Royco Photometer indicated that high dust concentrations were obtained only after a cleaning pulse.

When the collector was operated with no cleaning or intermittant cleaning (pulsing with process fan shut-off between filtering cycles) the mass efficiencies averaged >99.8% (99.26 to 99.98 wt% for 9 samples).

At one point the unit was operated with continuous cleaning (each bag pulsed alternately 4 times per min. per bag) using clean process air. An upstream sample indicated no fly ash present. The downstream sample, however, yielded dust loadings which were two thirds of those obtained under normal operation.

It seems likely that the fly ash is "seeping" through the felt. The cleaning pulse apparently stretches the felt allowing particles to work through toward the clean side of the cloth where they are resuspended. (Microscopically, the outlet dust looked very similar to the inlet dust.) This phenomena may be due to a peculiar interaction between the fly ash and polyester felt. To test this hypothesis, the bag, after being washed in mild soap and water, was used to filter talc. The mass efficiencies obtained were greater than 99.9 wt %.

SHAKER EXPERIMENTS

The intermittently cleaned shaker unit used in this experimental study was employed previously to filter gases from a sodium fire*. The unit was used essentially without modification (see Figure 26). One woven teflon bag (2/1 twill), 5 feet long x 5 inch diameter, was hung vertically in a square housing, 8 inch x 8 inch x 7 feet. The bag, a sleeve open at both ends, was clamped at the top and bottom to metal cylinders. Dirty gas enters at the bottom of the collector, into the top of dust hopper. The gas then flows into the bottom of the bags (a dust cake collecting on the inside of the bag) and leaves at the top of the collector.

The bag was cleaned intermittently by vertical shaking. The maximum number of shakes obtainable was about 230 shakes per minute.

The dust (fly ash), dust feeder, inlet and outlet samplers, and fan were the same as those used for the pulse-jet unit described above.

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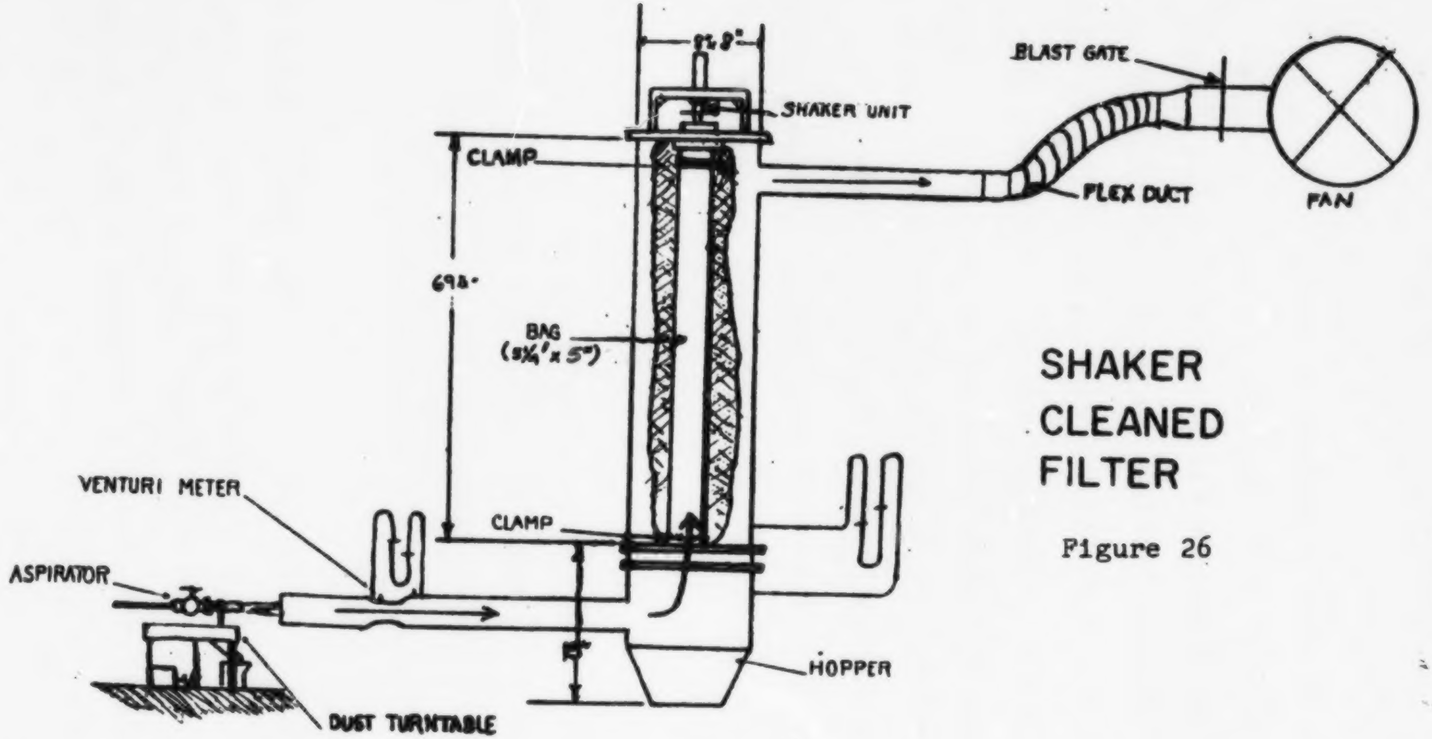
*See Semiannual Progress Report, Harvard Air Cleaning Laboratory, March 1 - Aug. 31, 1969 and Sept. 1, 1969 - Feb. 28, 1970.


9¼ 8"
SHAKER UNIT
BLAST GATE
FAN
FLEX DUCT
CLAMP
698·
BAG
(5¾' x 5")
CLAMP
VENTURI METER
ASPIRATOR
HOPPER
DUST TURNTABLE

SHAKER
CLEANED
FILTER

Figure 26

The shaker unit was tested on a gas containing about 6 grains fly ash per cubic foot air at an air to cloth ratio of 10:1 for 10 or 20 minute periods. It was then cleaned for 1 minute with about 200 shakes (occasionally longer with no detectable difference) and the dust was allowed to settle for 1 minute (sometimes longer). This procedure was repeated over a considerable length of time (11 hours). As can be seen from Figure 27, the pressure drop-time cycle did not stabilize. The reasons for this were believed to be one of the following:

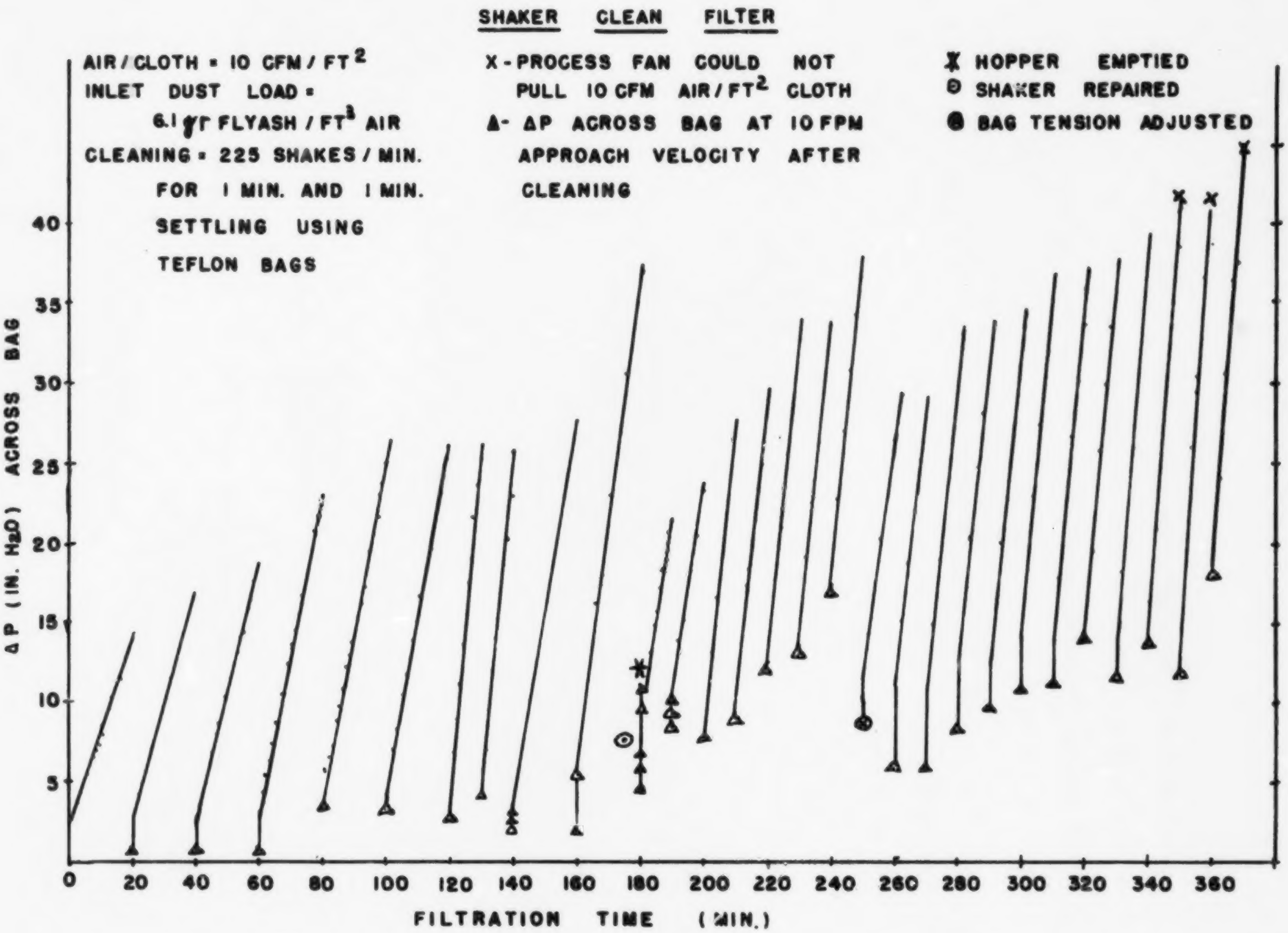
(1) The unit was originally designed for an air to cloth ratio of 2 CFM/FT^2. This is a typical air to cloth ratio for shaker cleaned filters.

(2) A Teflon bag was employed. This cloth is probably not a good support medium for a filter cake at high velocity since it is composed of smooth filaments.

(3) The dirty gas enters the dust hopper before entering the bag, possibly causing re-entrainment of dust as the hopper fills.

Installation of a cotton sateen bag yielded very little improvement. It was subsequently demonstrated that the design of the dust hopper was the primary cause of the problem. At 2:1 air to cloth ratio (design velocity) there is no noticeable

Figure 27


SHAKER CLEAN FILTER
AIR/CLOTH = 10 CFM/FT²
INLET DUST LOAD =
6.1 gr FLYASH/FT³ AIR
CLEANING = 225 SHAKES/MIN.
FOR 1 MIN. AND 1 MIN.
SETTLING USING
TEFLON BAGS
X - PROCESS FAN COULD NOT
PULL 10 CFM AIR/FT² CLOTH
Δ - ΔP ACROSS BAG AT 10 FPM
APPROACH VELOCITY AFTER
CLEANING
✱ HOPPER EMPTIED
⊙ SHAKER REPAIRED
⊗ BAG TENSION ADJUSTED
ΔP (IN. H₂O) ACROSS BAG
0
5
10
15
20
25
30
35
40
0
20
40
60
80
100
120
140
160
180
200
220
240
260
280
300
320
340
360
FILTRATION TIME (MIN.)

re-entrainment* of dust from the hopper. However, at 10:1 air to cloth ratio there is some re-entrainment, and at 20:1 air to cloth ratio, re-entrainment is so severe that operation of the unit is impossible.

The mass efficiency for the shaker unit was about the same for teflon and cotton sateen bags. The average mass efficiency was 99.95 wt% (99.8 to 99.99% for 7 samples.

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*The dust hopper was separated from the collector. Air was fed into the collector inlet (see Figure 26), and re-entrainment was measured qualitatively at the dust hopper exit using a Tyndall beam.

Section VI

BIBLIOGRAPHY

GENERAL

The following is a list of references with direct or peripheral relevance to the study of fabric filter systems. In compiling this information, emphasis was placed on material which is theoretical in nature.

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